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16. ABSTRACT The two-volume proceedings document presentations at EPA's Sixth Symposium on Flue Gas Desulfurization (FGD), October 28-31, 1980, in Houston, Texas. Presentations covered such subjects as approaches for control of acid rain, the Nation's energy future, economics of FGD, legislative/regulatory developments, FGD research/development trends, FGD system operating experience, FGD byproduct disposal/utilization, developments in dry FGD, and industrial boiler applications.

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Desulfurization	Regulations	Acid Rain	07A,07D	
Acidification	Byproducts		07B,07C	
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PREFACE

These proceedings for the symposium on "Flue Gas Desulfurization" constitute the final report submitted to the Industrial Environmental Research Laboratory, U.S. Environmental Protection Agency (IERL-EPA), Research Triangle Park, NC. The symposium was conducted at the Shamrock Hilton Hotel in Houston, TX, October 28-31, 1980.

This symposium was designed to provide a forum for the exchange of information, including recent technological and regulatory developments, on the application of FGD to utility and industrial boilers. The program included a Keynote Address on the approaches for control of acid rain, forecasts of energy and environmental technologies and economics for the 1980's, and sessions on the impact of recent legislation and regulations, research and development plans, utility applications, by-product utilization, dry scrubbing and industrial applications. Participants represented electric utilities, equipment and process suppliers, state environmental agencies, coal and petroleum suppliers, EPA and other Federal agencies.

Michael A. Maxwell, Chief, Emissions/Effluent Technology Branch, Utilities and Industrial Power Division, IERL-EPA, Research Triangle Park, NC, was General Chairman, and Julian W. Jones, a Senior Chemical Engineer in the same branch was Project Officer and Co-Chairman.

Franklin A. Ayer, Manager, Technology and Resource Management Department, Center for Technology Applications, Research Triangle Institute, Research Triangle Park, NC, was symposium coordinator and compiler of the proceedings.

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*Denotes speaker

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Session I: OPENING SESSION

Michael A. Maxwell, Chairman
Industrial Environmental Research Laboratory
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KEYNOTE ADDRESS

Approaches for Control of Acid Rain

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Assistant Administrator

Office of Research and Development

U.S. Environmental Protection Agency

Washington, D.C.

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The current situation in the Persian Gulf has once again brought home the stark reality of the fragile balance of our industrialized interdependent society. Once again we learn that our national economy can be tipped up or down by events thousands of miles away from our shores. National security and foreign policy deliberations must again focus on the question, "What are the likely impacts on our oil imports of a broadened war in the Mid-East?"

We have come to the point where we must find alternatives to foreign oil....and we have recognized that we have our own massive coal resources -- a wealth of "black gold" -- among the greatest known reserves existing anywhere in the world. We have recognized that we must move away from our dependence on foreign oil to greater reliance on domestic coal. President Carter and the Congress have mandated this conversion to coal as part of our overall National Energy Plan. We are beginning to move from a predominantly oil-based energy supply structure to one emphasizing domestic coal, oil shale, unconventional natural gas and heavy oil. And we are also encouraging -- and succeeding in -- a vigorous energy conservation program.

What this means, of course, is that we are going to be mining and burning more of the "dirtier" fuels. And that means there could be a growing air pollution problem. Coal mining in the U.S. is projected to increase from the current 700 million tons annually to 1.4 billion tons in 1990 and 1.9 billion tons in 2000. Conventional combustion will continue to be the primary method of utilizing this coal well into the twenty-first century -- despite the growth of a major coal-based synthetic fuel industry.

The challenge we face, therefore, is to maintain our air quality as the production of pollutants from burning fossil fuels rapidly expands. Because of the increased use of fossil fuels and the necessary cost of pollution abatement, there will be increasing pressure in the future to improve environmental control technologies, to make them more cost-effective and -- equally important -- to achieve widespread acceptance and operational utilization of these control systems by the utilities and industrial facilities. This 6th FGD Symposium is testimony to a continuing effort by both government and industry to meet these challenges.

The Congress has also provided impetus for the development and application of upgraded control technologies, like FGD. The 1977 Amendments to the Clean Air Act underscored the importance of control technologies through the requirement for Best Available Control Technology in areas where the air is clean....and the requirement for Lowest Achievable Emission Rate in "non-attainment" areas where the air is already dirty.

The recently issued New Source Performance Standards for utility boilers and the forthcoming development of NSPS for industrial boilers are typical examples of recent environmental protection efforts that will drive the continued research and development of environmental control technologies.

I think it is likely that Federal legislative action in the future will not significantly weaken current environmental programs. I believe, rather, that in the face of the pressures to relax environmental controls to allow more rapid expansion of our domestic fuels utilization, the public and Congress will continue the trend toward careful consideration of environmental impacts of future energy development. While we have made progress in improving air quality throughout the country over the last decade, the struggle is far from over. The recent smog episode in southern California is a grim reminder that some parts of the nation are still threatened with severe air pollution under poor meteorological conditions.

We have made great strides in developing and demonstrating highly efficient, reliable flue gas desulfurization technologies. While there are improved coal cleaning and new combustion technologies that are in the developmental stage, and some even at the demonstration and pilot test stages, FGD systems are currently the only viable sulfur control technology capable of general application over the next ten years. It has been estimated that by 1990, electrical utilities will have invested between $10 and $20 billion for construction and operation of FGD units.

I see from the program that Gerald McGlamery of TVA is going to discuss the economics of FGD systems a little later this morning. I'm sure that those of you here representing the utilities will be especially interested in what he has to say about TVA's latest cost studies and experience. From our own studies in this area, we believe that there is a good dollars and cents case for converting from oil to coal -- and that includes taking into consideration the use of FGD control equipment. Let me cite a few figures. To produce one million BTU's of heat, the cost of oil is $5.18, based on a price of $30 per barrel. To produce the same one million BTU's of heat, the cost of coal is $1.30, based on a price of $30 per ton. A power plant could save five cents per kilowatt-hour by making the conversion and using the best available scrubber, one with a 90 percent efficiency in reducing sulfur oxide emissions. This translates to a savings of $14 million per year for the average size electric generating plant being built today.

Where less stringent scrubber controls are required, savings could increase. According to conservative EPA projections for burning high-sulfur coals, a savings of 1/5 of a cent per kilowatt-hour would be realized by a utility that retires even a modern oil plant, writes off the investment, and replaces it with a new coal-fired facility outfitted with the best scrubber available.

In the United States, Japan and the Federal Republic of Germany, operating FGD systems using wet processes, such as lime or limestone scrubbers, continue to show improvement. Most of these processes are currently capable of removing well over ninety percent of the sulfur oxides in the flue gas. Here, in the U.S., lime and limestone scrubbers have been applied to coal with a wide range of sulfur content, and they have reliably removed the sulfur oxides from burning coals with one to four percent sulfur content. Many of these U.S. high sulfur coal FGD installations have operational reliabilities of over 90 percent. FGD installations on low sulfur coal have operational reliabilities of over 95 percent....which is similar to the Japanese experience with low sulfur coals.

One example of a key program in nonregenerable systems is the lime/limestone prototype test facility at TVA's Shawnee Steam Plant. You'll be hearing about the latest results from that operation during tomorrow morning's session. The results of this particular program are important because over 90 percent of the U.S. coal-fired electric generating capacity presently committed to FGD systems involves the use of similar lime/limestone processes. The Shawnee program has been directed toward obtaining answers to some of industry's concerns about long-term reliability of the process, the large quantities of waste sludge generated by the scrubber, and the high capital and operating costs involved. I believe major technological improvements and cost reductions are possible and will be realized, as we learn from programs such as this one.

FGD systems are now performing reliably and effectively both here and abroad. As I mentioned, in Japan, during the past decade FGD systems have been installed on a widespread basis. They have operated reliably and have had outstanding success in improving the air quality. Dr. Ando will speak on this subject in detail, but I'd like to cite a few statistics to demonstrate how these systems have proven themselves in Japan. There is no reason why they should not be just as effective here in the U.S.

Approximately 75% of the utility power generated in Japan is fossil-fired steam-electric. The balance is hydroelectric and nuclear powered. Of the fossil-fired capacity, 85% is oil-fired (most of the oil imported) and only 3% is coal-fired -- so you can see that their problem with foreign oil dependency is much worse than ours. But they have reduced sulfur oxide emissions from burning both oil and coal by 50% between 1970 and 1975.... and this has been due in great part to the use of FGD systems. They now have ambient SO_2 standards that are among the most stringent in the world -- about half the yearly average emission level that we allow.

Although Japan and the U.S. have both emerged as world leaders in developing and applying FGD systems, Japan has generally moved ahead more rapidly, because of its more serious commitment to solving its pollution problems. As of the beginning of last year, Japanese utilities had FGD systems installed, under construction, and planned for about 16% of their fossil-fired steam generating capacity....75% of it already installed and operating.

In the U.S., on the other hand, only about 3% of the total fossil-fired utility capacity is presently under FGD operational control. There are plans or systems under construction, however, for another 12% of the total fossil-fired capacity. At last count, 73 FGD units were in operation, with 127 units in design or under construction. When all of these units are operational, over 25% of the current total U.S. coal-fired capacity will be equipped with FGD. Because of this growing use of FGD, the total amount of sulfur oxides emitted to the atmosphere is expected to remain constant or even decrease slightly by the year 2000.

Even though we have made great strides in controlling sulfur oxides, we still have a long way to go to ensure that our expanded use of coal will not degrade the quality of our environment. EPA has been pursuing an aggressive air emissions program to control sulfur oxides, nitrogen oxides, and particulates -- all released from the burning of coal. And all contributors to a growing problem of acid deposition, more commonly referred to as acid rain. I am concerned that acid rain may become one of the most significant environmental problems of the coming decade. It already poses an environmental threat to our aquatic resources and possibly to our forest and agricultural resources as well -- a threat that could intensify with the full-scale development of our fossil fuel resources. We must therefore continue to work toward controlling the emission of not only sulfur dioxides, but also nitrogen oxides and particulates, before they get a chance to get out into the atmosphere and create acid rain problems.

Far from being a "gentle rain from heaven," acid rain can cause extensive ecological damage. In New York's Adirondack Mountains, for example, an area that was once a sport fisherman's paradise, acid rain has killed all of the fish in half of the high-altitude lakes. We cannot even guess at this time the extent of the damage in North American lakes, but we strongly suspect that tens of thousands of lakes are threatened, with millions of dollars in recreation benefits and commercial fishing at stake. Acid rain may also be playing a part in the decline in forest growth observed in both the Northeastern United States and southern Sweden. Experimental studies have shown that acid rain may damage foliage, interfere with the germination of seeds and the rooting of seedlings, affect the availability of nitrogen in the soil, decrease soil respiration, and deplete its nutrients. The destruction of stone monuments and statuary throughout the world, including the 2500 year-old Parthenon in Athens, Greece, has been accelerated by acid rain.

Acid rain may even indirectly present humans with a health hazard. If drinking water reservoirs become contaminated with acids, increases in heavy metal concentrations may exceed public health limits. In New York State, for example, water from the Hinckley Reservoir has acidified to such an extent that when the water comes in contact with household plumbing systems, lead from soldered joints passes into the water. These concentrations exceed the maximum levels recommended by the New York State Department of Health.

Acid rain was once thought to be primarily an SO_2 problem, but we've since learned that the phenomenon is more complicated than that. Nitrogen oxides as well as sulfur oxides can be transformed into potent acids when they combine with water vapor molecules in the atmosphere. The result is rain that may be -- as we have found in some parts of the country -- as acidic as lemon juice. Normal rainwater has a pH of about 5.7; newly hatched fish, which are most sensitive to low pH, are in serious trouble in water when its pH goes below 5.0. The average pH of the rain east of the Mississippi today is 4.4, which is almost 20 times as acidic as normal.

In the United States, the rain is most acidic in the heavily industrialized Northeast, but the most rapid increase in acid rain seems to be occurring in the Southeast. This parallels the expansion of Southeastern urban and industrial activities that result in sulfur and nitrogen emissions. Here, the trend is more apparent than in the Northeast, because the atmosphere is more rapidly deteriorating, and fewer acidic ions are required to cause a pH change. Most of the West has thus far escaped the acid rain scourge, but Colorado, the Los Angeles Basin, the San Francisco Bay Area, Spokane, Tucson, and Portland are known exceptions. In much of the West, the alkaline nature of the soils and lakes acts to neutralize acid rain, so the effects may not be as pronounced there. But even in the West, ominous signs of vegetation damage have appeared.

The Adirondack fish disaster, which occurred in an area of thin soils and fragile, closely watched ecologies, may be only a dramatic early warning of the damage that acid rain may someday cause on a much larger scale. Were it not for the buffering ability of the soil in other sections of the East Coast, the rains of the 1970's could have killed off most of the region's freshwater fish.

Clearly, we are not talking about something that sprang from the overactive imagination of a zealous environmentalist. Acid rain is a phenomenon that demands careful attention.

What can be done to prevent the rains of the 1980's from becoming increasingly more destructive? The most urgent task that EPA faces is to get to the bottom of what causes acid rain. Until the perplexing mechanisms by which acid rain is formed are better understood, attempts to control it may miss the mark, resulting in a less than optimum use of costly investments for control.

It _is_ known that, after sulfur and nitrogen oxides are discharged into the atmosphere, they are oxidized into sulfates and nitrates, which then react with moisture in the air to become acids. There are several complicated pathways or mechanisms by which this oxidation can occur. Which path is actually taken depends on a number of factors, including the concentration of heavy metals, the intensity of sunlight, the temperature, the humidity, the amount of ammonia present, and the particulate and photochemical smog levels.

In the eastern United States, sulfuric acid is the major component of acid rain, comprising as much as 65 to 70% of the rain's acidity, while nitric acid supplies only 25 to 30 percent. In the West, the acids in acid rain are generally half nitric acid and half sulfuric acid, although in some western urban areas, as much as 80% of a rain's acidity can be comprised of nitric acid. Other acids can also contribute to the acid rain problem. Hydrochloric acid, for example, may be emitted directly from coal-fired power plants and is frequently found relatively short distances downwind from such sources.

Acids may be deposited on earth not only by rain or snow, but also through an atmospheric process called "dry deposition." This is the process by which particles such as fly ash, as well as SO_2 and NO_x, are deposited onto surfaces. While these particles or gases are normally not in the acidic state before deposition, it is believed that they are converted into acids after contacting water in the form of rain, dew, fog, or mist after deposition. The precise mechanisms by which dry deposition takes place, and its effects on soils, forests, crops, and buildings, are not adequately understood. Much research is being initiated to clarify the contribution of dry deposition to the overall acid deposition problem.

Another aspect of acid rain that demands further study, and which makes regulation of acid rain a particularly tricky undertaking, is long-range transport. This phenomenon was first recognized in the early 1970's. At that time, studies on the adverse effects of SO_2 and sulfates on human health led to a stringent ambient air quality standard for SO_2 as well as technological control of SO_2 emissions. The associated control efforts forced the utilization of low sulfur fossil fuels and scrubbers, and resulted in lower sulfur dioxide emissions. Unexpectedly, however, reductions in urban SO_2 levels did not result in proportional decreases in urban sulfates.

Several theories were offered to explain this development. One explanation, the transformation-transport theory, was that reductions in urban SO_2 emissions were offset by increases in rural SO_2 emissions from new power plants located outside cities. SO_2 emissions from these power plants, the theory held, had been transformed into sulfates and transported over long distances to urban areas.

A project that was recently completed by EPA's Office of Research and Development on sulfur transformation and transport seems to bear this theory out. It found that sulfate aerosols could be transported hundreds of kilometers from the initial SO_2 source. This validation of the transformation-transport theory reinforces evidence indicating that the acidity of lakes in New York's Adirondack Mountains, for example, may be caused by acids carried by winds from power plants as far away as the Midwest.

Under certain conditions, it appears that sulfate and nitrate compounds can stay aloft long enough to cross continents, oceans, and international boundaries. This creates a situation in which the acid rain in one country is caused by the emissions of another, but the recipient of this damaging rain receives little or no benefit from the source initiating the pollution. In a few short days, local problems can become international in scope. This aspect of acid rain has caused us problems with our northern neighbor; Canada receives two to four times the amount of SO_x that the U.S. gets from Canada, and the NO_x exchange is 11 times greater from the United States to Canada. Recent negotiations between the two countries have been aimed at confronting this problem. These talks are expected to evolve into a bilateral transboundary air pollution agreement. And, through agencies like the United Nations Economic Commission for Europe, the acid rain issue vis-a-vis other countries may also be faced.

EPA is not alone in its efforts to uncover the causes of and the solutions to the acid rain dilemma. Many government agencies as well as private industry are participating in these efforts. In recognition of the seriousness of the acid rain threat, the President, in his Second Environmental Message, called for a minimum of $10 million per year to be spent over the next ten years on a comprehensive acid rain research program. He also established an Acid Rain Coordinating Committee consisting of seven Federal agencies to plan and coordinate a Federal interagency program. The Committee is co-chaired by representatives from the Department of Agriculture and EPA, and more recently, the National Oceanic and Atmospheric Administration. As one of the co-chairmen of the Federal Committee, I am pleased to note that the federal agencies are now spending over $15 million for acid rain research under the AEGIS of a cooperative research plan.

In addition to generating information on acid rain that can be used to develop air quality control strategies and options, EPA has another fundamental task: to communicate to Congress and the public the effects of acid rain, with particular attention paid to the ecologic and economic consequences of continued high levels of acid precipitation.

One tool to accomplish this communications function will be the development of an "acid deposition document," which David Hawkins, EPA's Assistant Administrator for Air, Noise and Radiation, and I are mapping out. This document will be an attempt to quantify and quality, in a preliminary way, the entire range of pollutants involved in acid rain creation -- sulfur, particulates, nitrogen oxides, hydrochloric acid, hydrocarbons and heavy metals.

The document will not be a "criteria document" in the sense that it will be used to develop ambient air standards; rather, it will put the scientific evidence on acid rain before the public so that it can be discussed in an open forum, as well as serve as a focal point for future acid rain research. This document, we hope, will be an important step toward fostering public debate about how we as a country will meet the acid rain challenge.

We do know, at this time, that some of the methods currently being used to minimize the local effects of SO_2 and NO_x around large sources are actually aggravating the acid rain problem. One method long favored by power companies is the use of tall emission stacks. The rationale behind tall stacks is that the emitted sulfur dioxide will be carried away from the local community by winds. Unfortunately, the tall stacks also keep the sulfur dioxide airborne longer, thus making sulfate formation more likely.

As the mist that conceals the secrets of acid rain formation and transport is gradually lifted, we will know better what control methods will actually stop acid rain at its source, rather than passing the problem on to someone else. At present, however, it appears that the only practical approach lies in reducing SO_x and NO_x emissions. Many innovative schemes have been suggested. There are studies underway to estimate the costs of various ways to reduce emissions of these pollutants and to compare these costs against acid rain damage costs, which are only now beginning to be understood.

For SO_x control, FGD will probably remain our chief weapon through at least 1985. As you will hear throughout this conference, this technology can be applied to a variety of sources without imposing an unreasonable financial burden. The use of low-sulfur coal is another piece of the arsenal in the war against SO_x emissions, along with the array of technologies, both under development and on the commercial market, designed to remove sulfur from fuel before it is burned. These technologies include coal cleaning, coal gasification, and desulfurization of liquid fuels. Then, there are also the combustion modification methods that allow removal of sulfur during burning, such as fluidized-bed combustion.

But, as we have seen, SO_x constitutes only a piece of the acid rain puzzle. NO_x emissions can play an equally large role. And while we have found ways to hold the lid on SO_x emissions, we've only recently begun to get a handle on NO_x control. In fact, as coal use rises, we expect that NO_x emissions could increase by thirty to forty percent by the year 2000, unless more effective control methods are developed and quickly put to work by industry. At present, half the current NO_x emissions come from stationary sources; but by 2000, due to the trend toward greater combustion of coal, stationary sources may be responsible for up to 75 percent. Of the emissions from stationary sources, over half are contributed by utility and large industrial boilers alone. These large boilers now emit an estimated 6 million tons of NO_x every year.

The solutions that are so effective for SO_x control aren't much help when it comes to NO_x control. Physical coal cleaning, which can be used on some coal to reduce sulfur and ash content, has no effect on coal's nitrogen content, because the nitrogen is chemically bound to the coal. "Denitrogenation" -- that is, chemically removing nitrogen from coal -- is prohibitively expensive at present, and at any rate does not address the problem of thermal NO_x, which is formed by molecular reaction in super-heated combustion air. Flue gas treatment for NO_x control has been used with a fair amount of success in Japan on oil-fired boilers, but there are major financial and technical hurdles to applying that technology to coal-fired units. Even the coming age of synthetic liquid fuels made from coal, which may consume 120 million tons of coal in 1990 and 300 million tons in 2000, offers little hope for NO_x control -- in fact, the concentration of fuel nitrogen may be increased when coal is converted to a liquid.

However, there is a promising answer that is both cost-effective and energy-efficient. By modifying the conditions under which combustion takes place, an existing coal-fired power plant can reduce its NO_x emissions by 40 to 50 percent. When applied to new burner designs, combustion modification may reduce NO_x emissions by another two-thirds, yielding a total NO_x control of up to 85 percent. And, because combustion modification involves changes only in burner design, the cost is quite small -- less than one-half of one percent of the boiler cost. Further, because we are ensuring that the new burners are as efficient as the older designs, the operating cost is nearly zero. EPA is aggressively developing low-NO_x burner designs.

Ideally, one technology would simultaneously control both of acid rain's major components. This, in fact, is the idea behind a particularly exciting new control technology, which may be retrofitted to many existing coal-fired boilers with only minor modifications: the limestone injection/multi-stage burner, or LIMB for short. The LIMB may be able to remove 50 to 70 percent of sulfur oxides at the same time that it reduces NO_x by 50 to 80 percent. And it can accomplish this at a cost for SO_2 control equipment of only $30 to $40 per kilowatt, as opposed to the average of $150 per kilowatt that wet scrubbing requires.

Although the LIMB has only reached the bench/pilot scale stage of development here in the U.S., Germany is currently operating a 60 megawatt electric boiler using the technology, so we know that it works on a larger scale.

The idea of combining limestone injection for SO_2 control with a low NO_x burner is not a new one. In 1967, UOP, building on earlier limestone injection experiments by Combustion Engineering, injected limestone into an arch-fired burner, which is a naturally low NO_x burner. SO_2 emissions were reduced by 50 percent at a stoichiometric ratio of 1:3.

The 60 megawatt prototype limestone injection boiler in Germany, which I mentioned earlier, has been operating for one year. It fires West German lignite, and utilizes flue gas recirculation to minimize peak temperature and NO_x formation. At present, it is achieving 50 to 90 percent SO_2 removal at stoichiometric ratios of 2.5 to 5.0. Retrofit capital costs for this technology are only $3.00 per kilowatt.

EPA has proposed a five-year research, development, and demonstration program that will bring the LIMB technology up to commercial scale. In the first year, EPA will characterize reactions and furnace conditions; evaluate impacts on furnace operation; and test the technology with a wide range of coal types and calcium-based sorbents. Next will come a year of field evaluation, in which EPA goals will be to demonstrate sulfur removal efficiency, optimize performance variables, determine if there are any adverse boiler side effects such as slagging, plugging and corrosion, and obtain design and cost data. Both wall-fired and tangentially-fired units will undergo testing. Another year will be spent installing the LIMB technology on full-sized boilers, which will then be subjected to two years of performance optimization and long-term evaluation. The development effort will be co-sponsored by EPA and the Department of Energy. The total tab for the LIMB program will amount to $16.5 million, which will be a bargain if LIMB fulfills its initial promise.

Industry as well as government must play a crucial role in the development of methods to control acid rain. EPA has the resources to provide the fundamental research and the testing of new control technologies, but we must rely on industry to provide the host sites that allow technologies to be tested under real-life conditions. And, we must depend heavily upon the commercial expertise and engineering experience of boiler manufacturers if a technology is to progress beyond the demonstration stage.

Now there's always an element of risk for the private sector when it invests in new equipment and new technologies. Control processes that look promising on the drawing board or during small-scale experiments don't always pan out when they are put into practical use. But we at EPA truly believe that with the kind of cooperation between government and industry we have enjoyed up to now, and with continued joint effort, we can solve the acid rain control challenges we face.

With a better understanding of what causes acid rain and with the necessary control technology under development, we will be able to begin making strides in the regulatory arena....to pull in the "reins," if you will forgive me, on acid rain. As the Clean Air Act stands now, there are no regulatory requirements concerning acid rain *per se*. As most of you are aware, this Act comes up for revision next year, and EPA is consulting with other Federal agencies on the possibility of changes that would better address the acid rain issue.

The Clean Air Act is currently structured around a presumption that air pollution can be related to a particular source or a well-defined group of sources. But, in the case of acid rain, there is no clear-cut relationship between specific emissions and the acid rain. In other words, even though the types of emissions that lead to acid rain are known, it is currently not possible to accurately trace individual emissions that cause acid rain back to their origins. And, while the Clean Air Act has been amended to address the problem of interstate pollution, any given state is only able to enforce its emission limitations against sources within its own boundaries. A state can petition the EPA Administrator if it feels that another state is preventing it from attaining a national standard or otherwise causing a deterioration in that state's air quality, but then EPA is faced with the problem of how to demonstrate that one or several out-of-state sources are responsible for impermissible air quality violations. Such a demonstration would be hard, if not impossible, to make, especially if a number of sources from several states or nations were involved.

One regulatory option that EPA is reviewing is the development of national ambient air quality standards for nitrates or sulfates, two precursors of acid rain. However, it is not clear whether there is sufficient data on which to base such a standard. Even if the data were available, the standard-setting process is a lengthy one. It would probably be five to ten years before any emission reduction could be achieved. Other near-term options include: better monitoring of SO_2 emissions to improve enforcement of existing standards; the establishment of federal regulatory requirements for review of interstate impacts of State Implementation Plan provisions; or the establishment of new source performance standards for pollutants for which EPA has not set ambient standards, such as total sulfur.

A longer term option might involve the Congress setting regional SO_2 and NO_x emission reduction goals -- say 5 to 10 percent per year -- goals which would be administered on a multi-state basis and would allow the utilities and industries to meet the goals on a system-wide basis using the most cost-effective combination of approaches -- coal washing, combustion modification, load shifting to cleaner plants, fuel shifting, and early plant retirements, to name a few.

Whatever path we choose, however, we must be mindful of the need to consider the regulatory burden imposed on the utility or industry and the ratepayer or consumer. In addition, we must fully support the national energy policy of expanded coal use, and be sensitive to the fact that the economy cannot regain its vital growth without the aid of a vigorous industrial base. These are "mighty tall" orders, as they say, for the Government and the industrial sector. But then few people really believe that anything worth doing in this country is going to be easy. Why should reconciling environmental and energy goals, a priori, be any easier than, say, reconciling, energy goals and national security, or inflation and unemployment objectives. There are no easy answers, only a nation of differing but robust people trying to work out their future.

THE NATION'S ENERGY FUTURE - WITH FOCUS ON SYNFUELS

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ABSTRACT

Projections indicate that coal, nuclear energy and oil shale will become increasingly important as we adjust for static domestic oil and gas production and minimization oil importation. Environmental problems can be quite severe for each of these fuel cycles. A massive synthetic fuel industry based on coal, oil shale and biomass, is emerging with monumental potential for environmental damage. The Environmental Protection Agency (EPA) has designed a regulatory program aimed at mitigating environmental damage while allowing for birth and nurturing of this critical industry.

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THE NATION'S ENERGY FUTURE - WITH FOCUS ON SYNFUELS

OUR ENERGY FUTURE

America is making progress in minimizing dependence of imported oil. During the first five months of 1980, gasoline consumption decreased 8.1 percent -- compared with the same period last year - and crude oil imports decreased to 7.8 million barrels per day -- the lowest level in four years. Petroleum stockpiles are at capacity levels nationwide due to a very real, conscientious effort to conserve energy in all areas: electricity, home heat and transportation fuel.

Of even greater significance is passage of the Energy Security Act, signed by President Carter in June of 1980. This bill will promote conservation, increase production of coal and oil, and help harness the power of the sun, wind and rivers and most importantly spawn a major synthetic fuel industry based on coal, oil shale and biomass. All of these measures can serve as effective remedies against further reliance on costly and uncertain supplies of foreign oil.

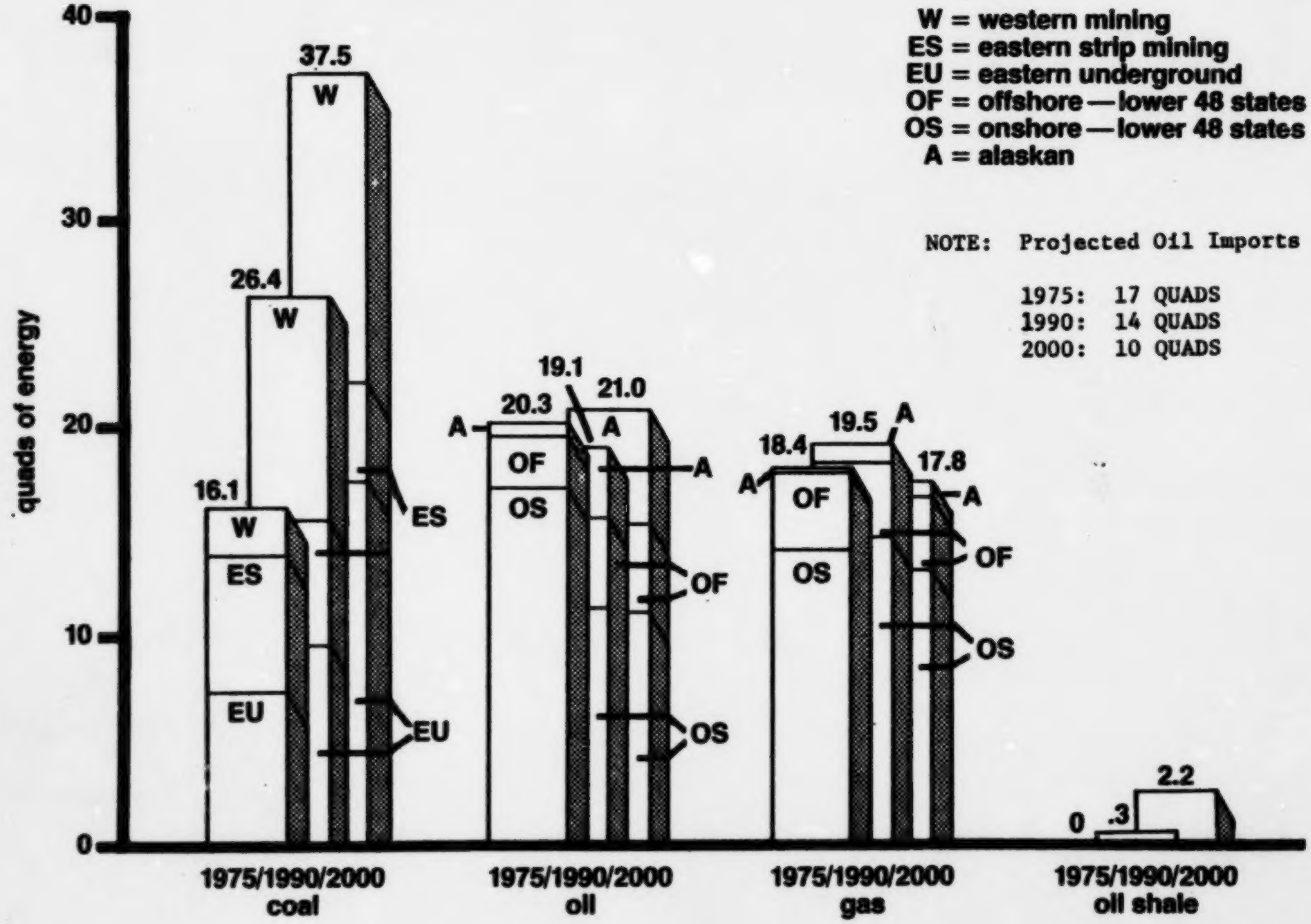
To achieve the necessary growth in domestic energy resource development to meet our future production goals, a substantial increase in extraction, processing, transport and use of domestic fossil fuels must take place. EPA has recently made projections attempting to predict our nation's energy future using the Strategic Environmental Assessment System (SEAS) model and an EPA sponsored study projected synfuel production. These projections suggest that coal, oil shale and nuclear energy will allow for the nation's economic growth despite the leveling off of domestic petroleum and natural gas and without increasing oil imports (Figure 1-4). For example, the amount of coal mined in this country must expand from the current 700 million tons annually to 1.1 billion tons in 1990 to 1.6 billion tons in 2000. The production of synthetic liquid fuel and gas from coal is expected to consume 80 million tons by 1990 and 350 million tons in 2000. We can also expect that the 1980's will see the oil shale industry emerge as a significant supplier of fuel, producing up to 300,000 barrels per day by 1990 and 2.2 million barrels per day by 2000.

Such projections indicate a trend away from traditional and less environmentally damaging energy sources, toward potentially more damaging fossil fuel sources such as coal (particularly from western surface mines), oil or gas from the Outer Continental Shelf, and western oil shale. The trend also points to the increasing use of nuclear energy to generate electricity and indicates an increasing interest and use of solar and geothermal energy.

These major shifts toward increased use of less clean fuels can pose a significant threat to human health and the environment. Potential negative impacts are likely to result from the extraction, processing and utilization phases of each major fuel (Figure 5). For example, increases in coal and oil shale mining can create erosion and subsequent surface water siltation problems:

FIGURE 1

domestic fossil energy resource requirements



NOTE: Projected Oil Imports

1975: 17 QUADS
1990: 14 QUADS
2000: 10 QUADS

FIGURE 2

U.S. non-fossil energy resource requirements

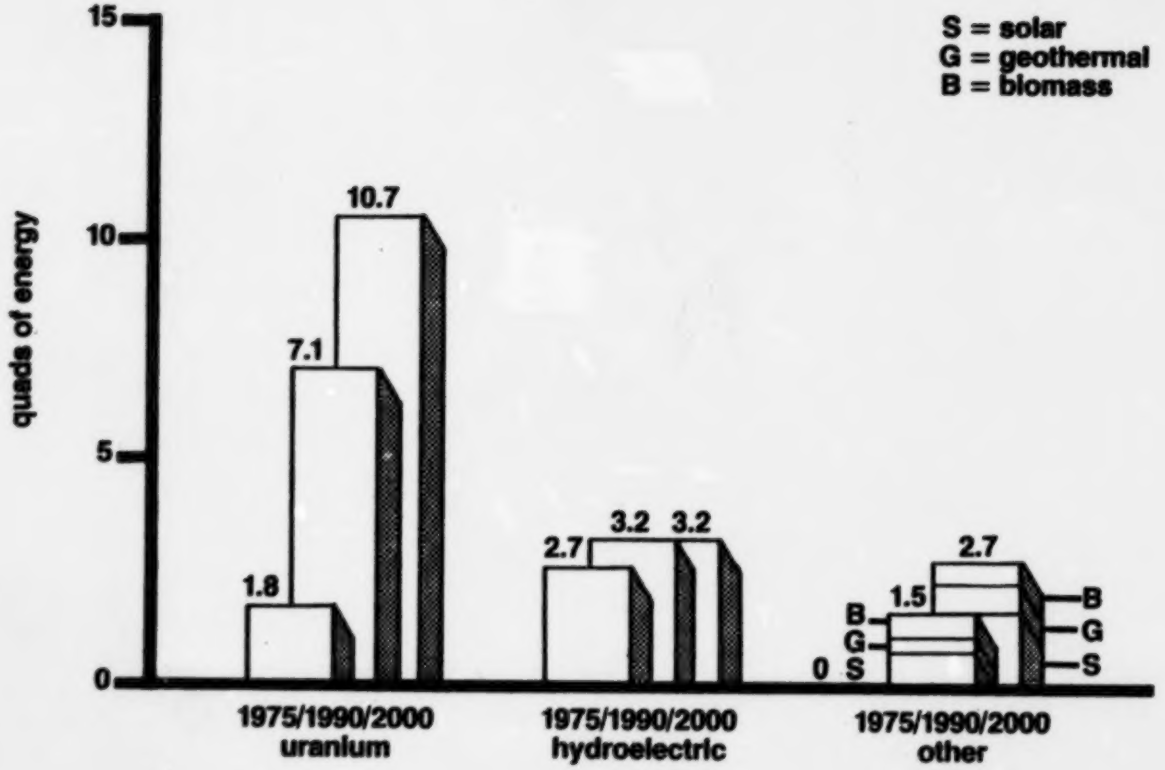


FIGURE 3

components of domestic coal utilization

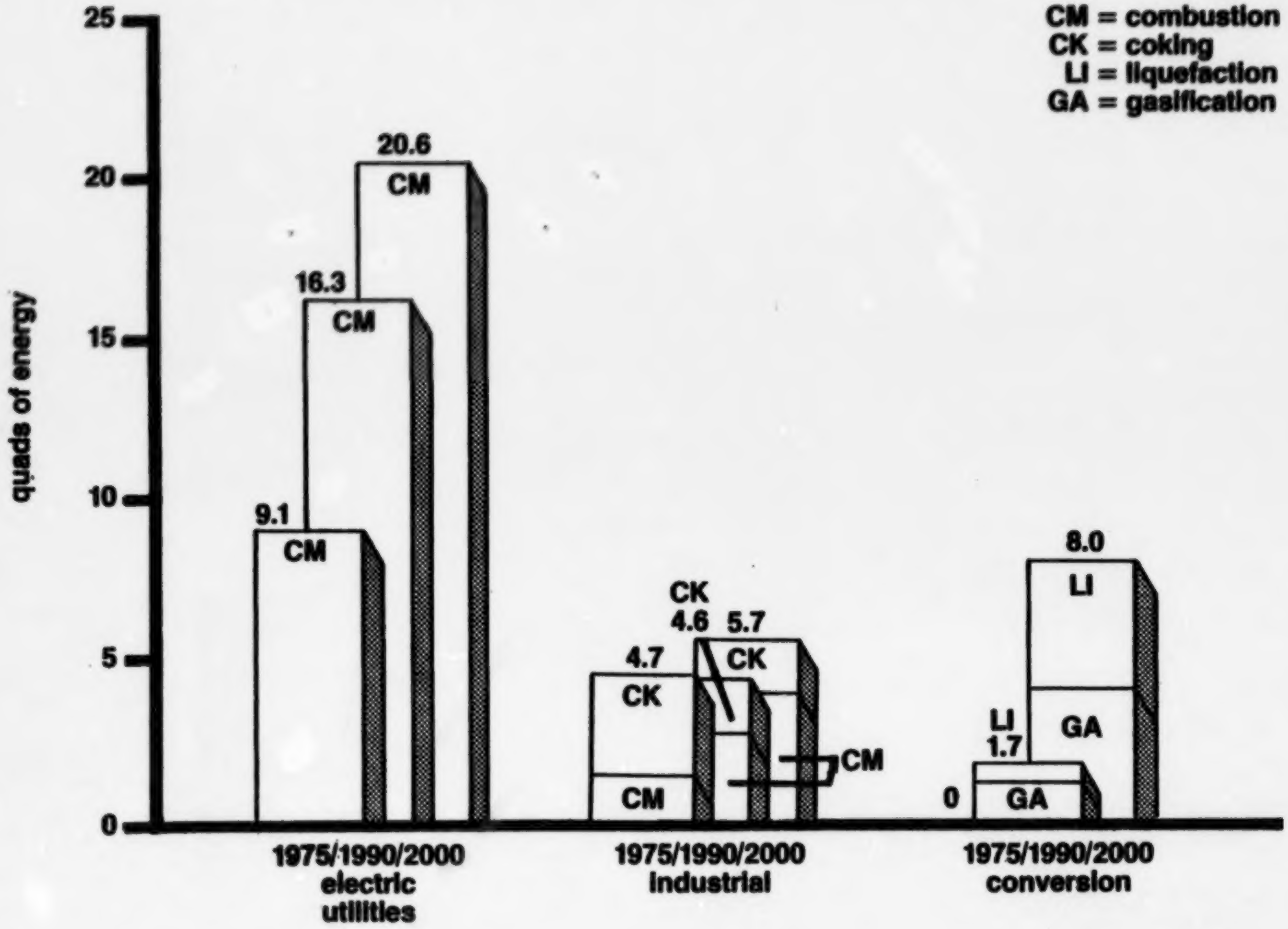
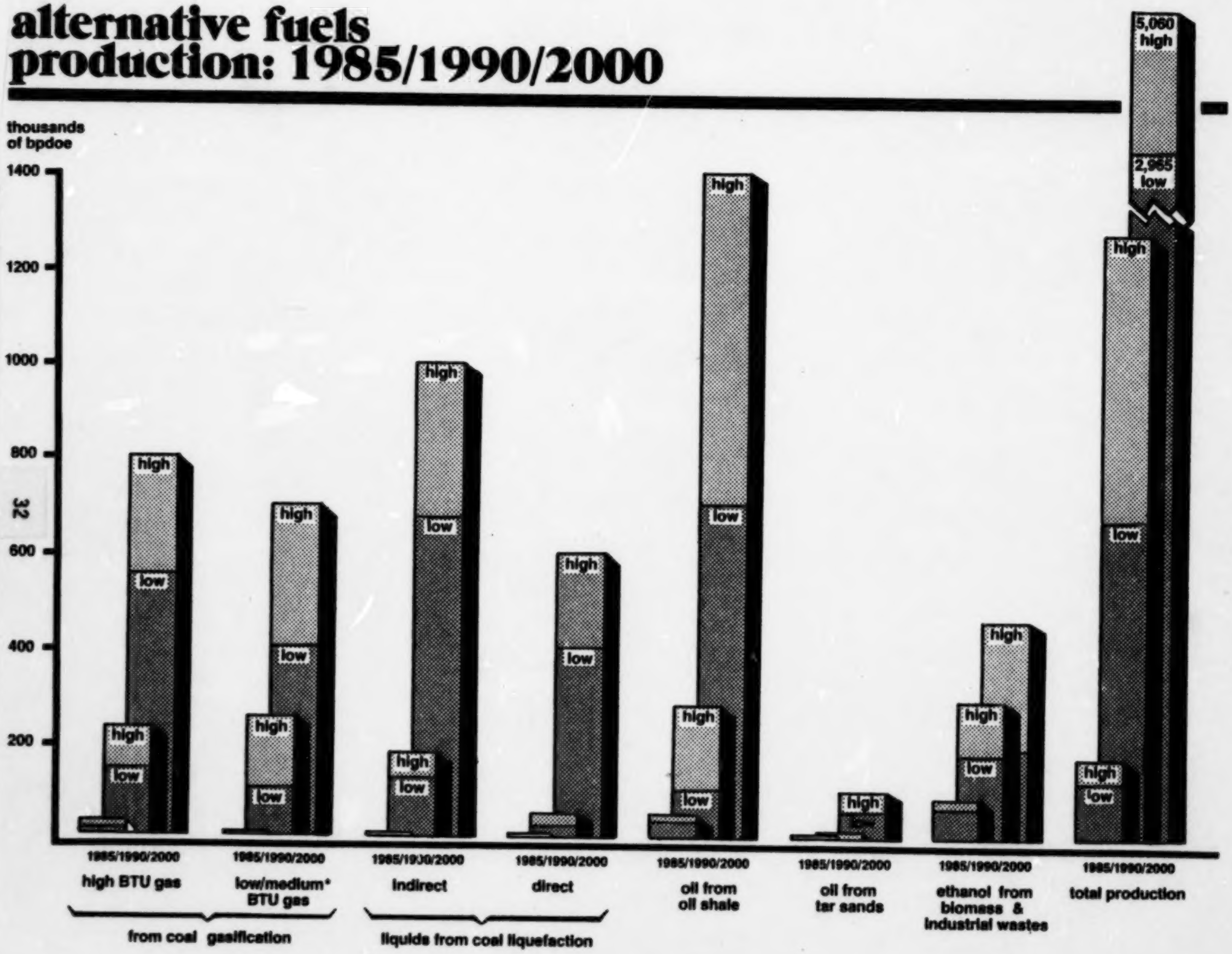


FIGURE 4

alternative fuels production: 1985/1990/2000

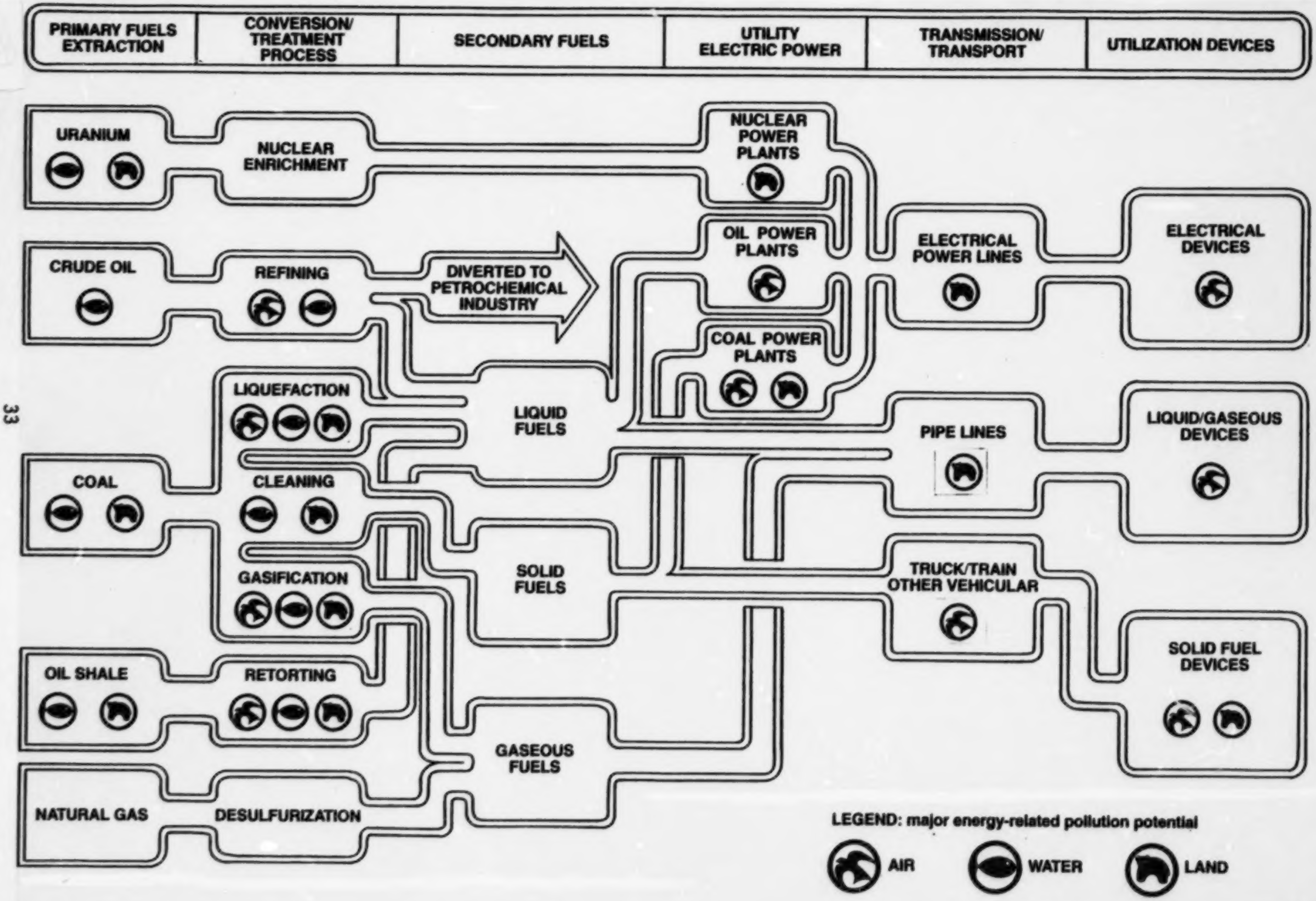


*predominantly medium

Source: Hagler, Bailly & Company *Alternative Fuels Monitors: Coal Gasification and Indirect Liquefaction; Oil from Shale;* Hagler, Bailly & Company estimates based on projections by Department of Energy; General Accounting Office; Booz-Allen; Office of Technology Assessment; House Committee on Interstate & Foreign Commerce; EXXON; ESCOE; MITRE; SERI; ERAB.

IGURE 5

simplified U.S. domestic fuel flow: 1990


PRIMARY FUELS EXTRACTION
CONVERSION/ TREATMENT PROCESS
SECONDARY FUELS
UTILITY ELECTRIC POWER
TRANSMISSION/ TRANSPORT
UTILIZATION DEVICES
URANIUM
NUCLEAR ENRICHMENT
NUCLEAR POWER PLANTS
CRUDE OIL
REFINING
DIVERTED TO PETROCHEMICAL INDUSTRY
OIL POWER PLANTS
COAL POWER PLANTS
ELECTRICAL POWER LINES
ELECTRICAL DEVICES
LIQUEFACTION
LIQUID FUELS
PIPE LINES
LIQUID/GASEOUS DEVICES
COAL
CLEANING
SOLID FUELS
GASIFICATION
TRUCK/TRAIN OTHER VEHICULAR
SOLID FUEL DEVICES
OIL SHALE
RETORTING
GASEOUS FUELS
NATURAL GAS
DESULFURIZATION
LEGEND: major energy-related pollution potential
AIR
WATER
LAND

groundwater quantity and quality are also likely to be affected. Processing coal and oil shale to synthetic liquids and gases may yield toxic emissions and large quantities of solid wastes; and despite current regulations, an increase in coal combustion will result in increased production of nitrogen oxides, sulfur oxides and solid wastes (Figure 6). The environmental and safety uncertainties surrounding the use of nuclear energy have been well publicized.

Many of the adverse impacts on health and environmental quality, however, can be controlled or avoided: Most mined land can be reclaimed; particulate matter and the oxides of nitrogen and sulfur can be scrubbed from flue gas; acid precipitation and its effects on agricultural and forest production can be reduced.

EPA has an impressive array of legislative tools available to control air, water and land pollution from energy and industrial sources (Table 1). The agency will face the monumental challenge of utilizing these mandates to achieve maximum benefit of minimum cost.

Controlling these pollutants increases the monetary costs of energy, but failure to control them lowers the productivity of our natural resources, degrades the quality of our environment, and imperils the health of our population.

Focus on Synthetic Fuels

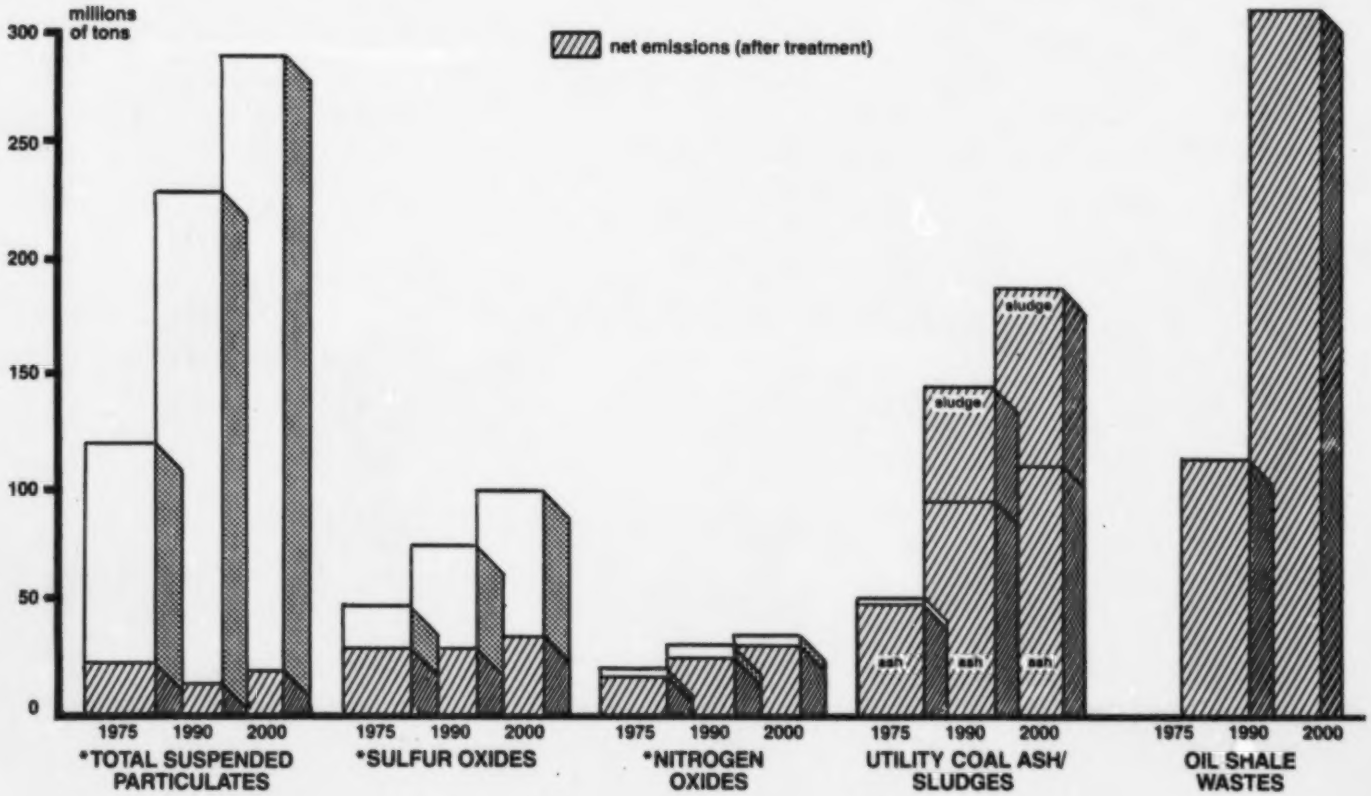
As the projections suggest our energy future should be characterized by a massive synthetic fuel industry by the year 2000. Although oil shale plants will be limited to a relatively limited area (Figure 7) coal gasification and liquefaction plants could be constructed anywhere large quantities of coal are located (Figure 8). Ethanol plants will be initially sited in corn and wheat farming areas (Figure 9) but could eventually proliferate as other crops and agricultural wastes become feasible as feedstocks (Figure 10).

oo Synfuel Environmental Issues

Synthetic fuels processes are receiving our most serious attention because synfuel development activity is clearly intensifying, because of our concern over the unknown nature of the pollutants which may be generated, and because of EPA's recognition that the enormous capital outlays involved in building these facilities during the next decade dictates the earliest possible and most stable possible environmental regulations for this new industry. It is expected that pollutants coming from coal conversion and shale oil production will be more diverse in composition than those produced by direct fossil fuel combustion. The burning of fossil fuels in conventional processes involves complete oxidation (or attempts threat) whereas synthetic fuels are produced under

FIGURE 6

growth in emissions/wastes from stationary sources



*energy and industrial sources

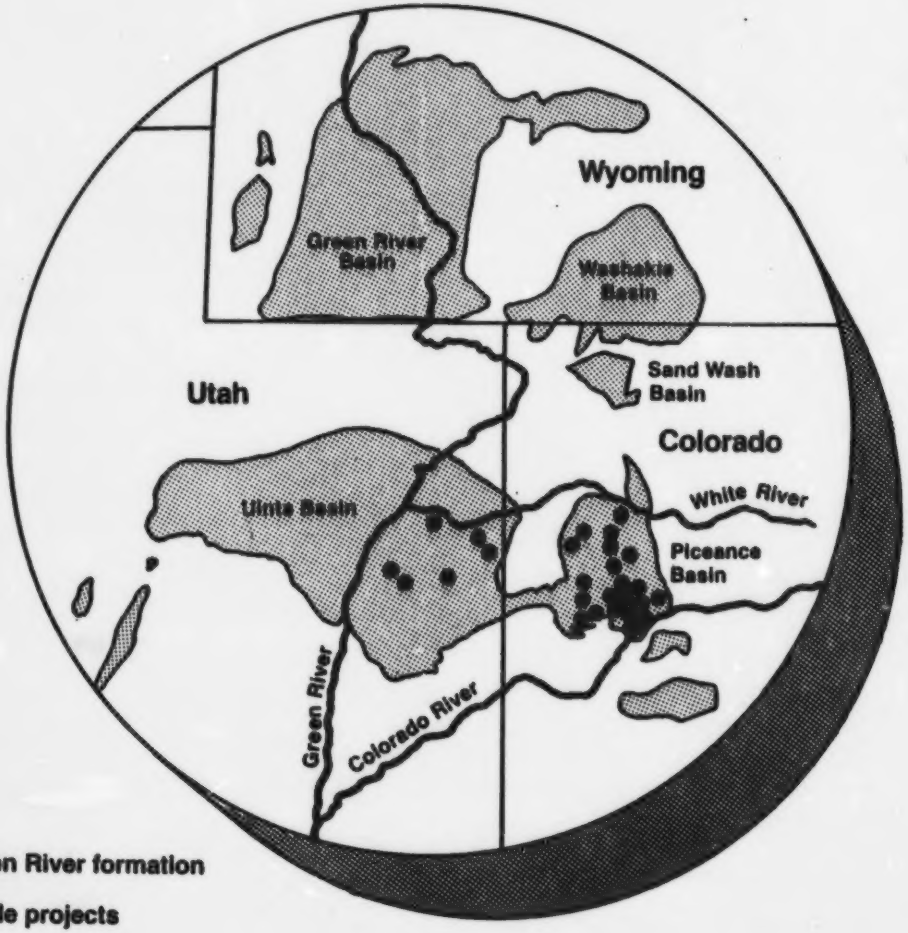
TABLE 1

AIR, WATER AND SOLID WASTE ENVIRONMENTAL LAWS IMPACTING FOSSIL ENERGY FACILITIES

RELEVANT AUTHORITY	IMPACT
Clean Air Act Amendments of 1977	
● Set New Source Performance Standards (NSPS) for energy industries (Section 111).	● NSPS set for fossil utility boilers; industrial boiler NSPS being developed; oil shale, coal gasification, and liquefaction in planning stage.
● Set National Emission Standards for Hazardous Air Pollutants (NESHAP) for selected industries (Section 112).	● NESHAP requirements for synthetic fuels industry being evaluated as process plans become firm.
● Implement Prevention of Significant Deterioration (PSD) Program (Section 160).	● PSD permits required for all New Sources (coal-fired boilers and synthetic fuels plants) to prevent increases in particulate and SO_2 levels in areas having good air quality.
● Achieve Ambient Air Quality Standards (Section 109).	● Require utilization of appropriate control technology to reduce emissions to levels required to meet State Implementation Plan (SIP) goals.
● Set Lowest Achievable Emission Rates (LAER) (Section 171).	● Require level of pollution control technology greater than that which would normally be required by SIP for plant siting in non-attaiment areas.
Federal Water Pollution Control Act Amendments of 1977	
● Set discharge limits based on best conventional technology for energy industries (Section 306).	● Effluent guidelines for steam-electric industry issued, industrial boilers must meet guidelines for specific industry; effluent guidelines being planned for oil shale and coal gasification and liquefaction facilities.
● Set discharge limits based on best available technology for toxic pollutants (Section 307).	● For designated toxic pollutants best available control technology will be required, and will have greatest impact on the design of synfuel plants.
● Issue and enforce discharge permits to achieve above limits and to meet water quality standards (Section 402).	● Permits for electric utility plants and other industries being issued based on effluent guidelines; permits for synthetic fuels plants will be issued on basis of best information available until guidelines are issued.
Safe Drinking Water Act of 1974	
● Review projects for possible danger to underground drinking water supplies (Section 1424).	● All projects receiving federal assistance will be reviewed for processes impact on groundwater quality as it may impact drinking water.
Resource Conservation and Recovery Act of 1976	
● Set criteria for defining hazardous waste (Section 3001).	● Proposed procedures for determining if wastes are hazardous have been issued.
● Define acceptable disposal practices for hazardous wastes (section 3008).	● Utility wastes and spent oil shale classified as "special" wastes; if hazardous, they must meet monitoring requirements but not disposal requirements; best economically attainable disposable technology will be defined.
● Set guidelines for non-hazardous waste disposal (Section 4004).	● Disposal guidelines for non-hazardous utility waste will be completed in 1981, other energy wastes subject to state guidelines.

FIGURE 7

some oil shale development sites


Wyoming
Green River Basin
Washakie Basin
Sand Wash Basin
Utah
Colorado
Uinta Basin
White River
Piceance Basin
Green River
Colorado River

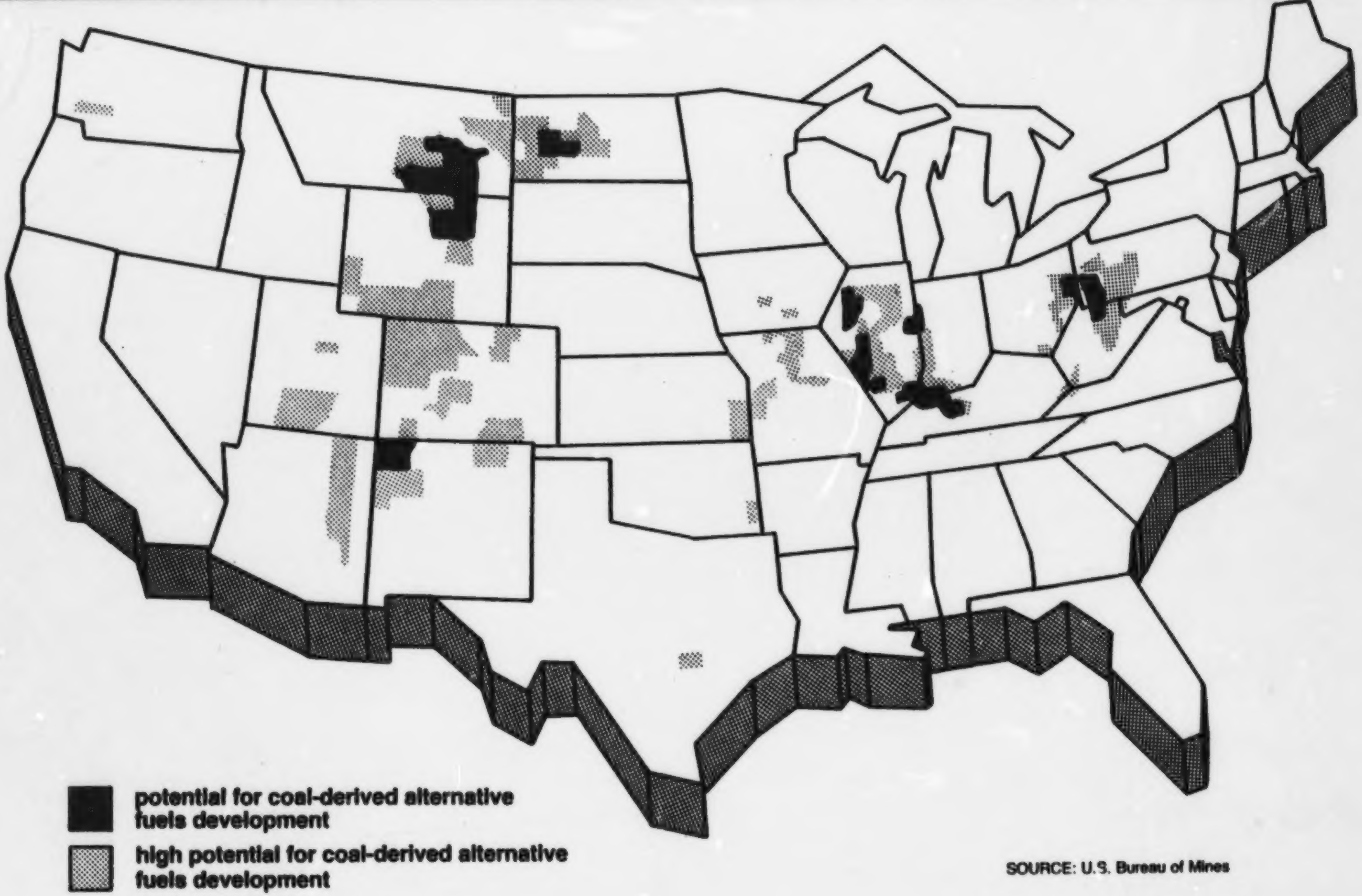
▨ oil shale deposits of the Green River formation

● planned and potential oil shale projects

SOURCE: Office of Technology Assessment

FIGURE 8

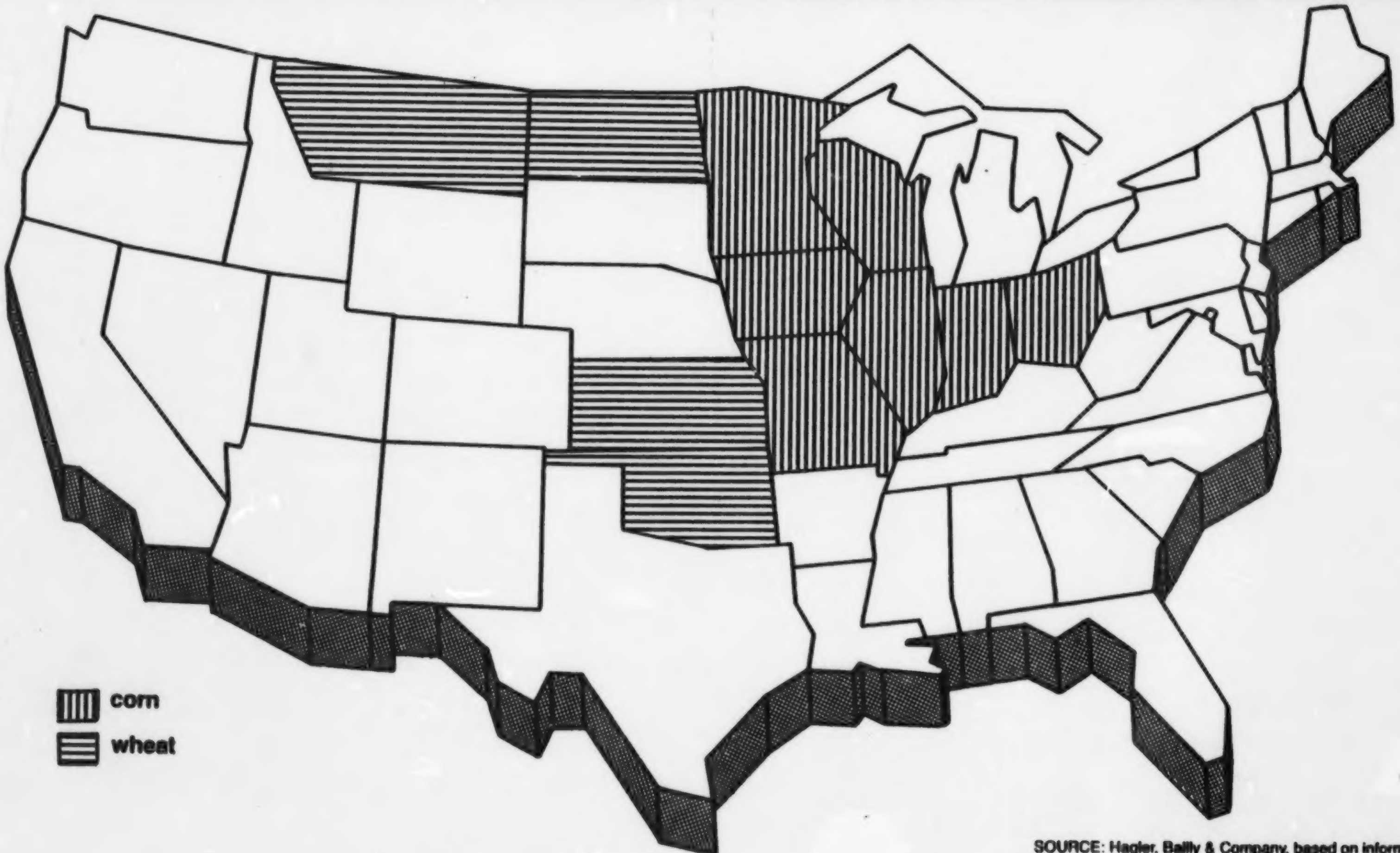
sites for coal-derived alternative fuels plants


■ potential for coal-derived alternative fuels development
▨ high potential for coal-derived alternative fuels development

SOURCE: U.S. Bureau of Mines

FIGURE 9

potential sites for large-scale ethanol production by 1985

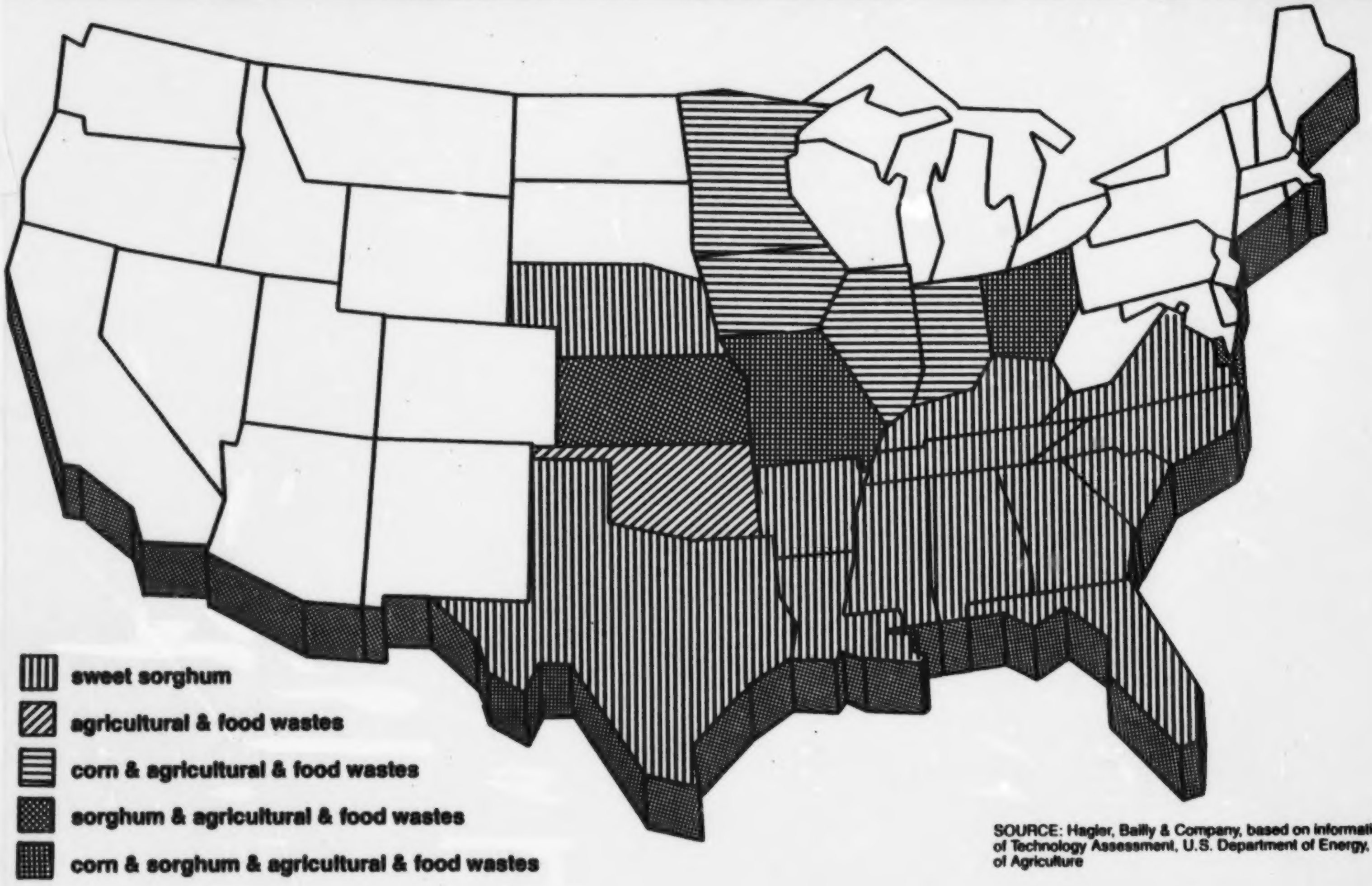


SOURCE: Hagler, Bailly & Company, based on information from Office of Technology Assessment, U.S. Department of Energy, U.S. Department of Agriculture

9590

FIGURE 10

potential sites for large-scale ethanol production by 2000


▥ sweet sorghum
▨ agricultural & food wastes
▤ corn & agricultural & food wastes
▩ sorghum & agricultural & food wastes
▦ corn & sorghum & agricultural & food wastes

SOURCE: Hagler, Bailly & Company, based on information from Office of Technology Assessment, U.S. Department of Energy, U.S. Department of Agriculture reducing conditions using less air than is required for complete combustion. The result is that a wide variety of high molecular weight organics, reduced sulfur compounds, and other potentially toxic compounds are formed, presenting a different array of pollutants than have been dealt with in the past.

We believe the air pollution problems may be particularly serious. The synthetic fuel industry is expected to produce a wide range of air emissions with potentially adverse environmental effects if not adequately controlled. Oil shale retorting, for example, will emit nitrogen oxides, sulfur oxides, reduced sulfur species, ammonia, various volatile and partially oxidized organics and, of course, particulate matter. The Prevention of Significant Deterioration increments available may well pose serious problems. The air pollution problems associated with coal gasification and liquefaction are similar in many ways to those for oil shale. These processes can generate significant quantities of particulates, sulfur compounds, trace metals, high molecular weight hydrocarbons and nitrogen oxides, etc. The sulfur species may be particularly troublesome.

Water-related environmental problems from synfuel production may be just as complex. The oil shale industry will need copious water supplies for cooling compaction of spent shale, and for revegetation of surface mined areas. Coal mining and coal conversion will also have substantial water requirements for process uses and revegetation. Supply of water for these activities will be particularly crucial at some sites in the arid western part of the country where oil shale retorting and some mine-mouth coal conversion will occur. At other sites, mine dewatering and retort-produced water from shale oil production will produce excess water. Among the water pollution problems of concern, spent shale, if not properly handled, could create serious water quality problems from the leaching of soluble contaminants into nearby ground or surface water. With underground, modified "insitu" operations being considered for oil shale, and possibly for coal, the opportunity for groundwater contamination is even more likely than for surface operations. Here again, the problem is particularly serious in the western part of the country where groundwater is a vital resource. From all types of synthetic fuel operations, raw process water discharges will be highly contaminated by toxic materials (most likely including carcinogens, mutagens, etc.) which would represent major threats to both surface and groundwaters if not properly controlled. It is expected that synfuel facilities will utilize process water recycling to a great extent but this may not totally solve the water pollution problems at all locations.

There are a variety of synfuel-related solid waste problems as well. Both oil shale mining and coal mining produce enormous amounts of solid waste. Many of the mining problems are similar to those encountered with conventional coal mining and can be solved similarly. Surface reclamation techniques for strip mined areas are particularly successful at least where an adequate water budget exists. The solid residues of oil shale retorting and coal conversion are, however, another problem. Shale oil production, for example, produces spent shale that is greater in volume than the shale originally removed from the ground; coal conversion technologies, both gasification and direct liquefaction, will produce vast quantities of ash. Each of these wastes will most likely contain a wide variety of potentially harmful components and will have to be properly managed. Some special wastes from synfuel plants such as spent catalyst from coal conversion may be classified as "hazardous" under the Resource Conservation and Recovery Act.

There is also concern about the possible toxicity of liquid synthetic fuels themselves, both from the handling and usage standpoints, including concern for both industrial employees and the general public. Coal-derived liquid fuels, particularly those produced by direct liquefaction, are of the most concern. These liquid fuels are not of the same composition as ordinary crude oil products. They are higher in nitrogen content, yielding higher NO_x levels upon combustion and they tend to contain more substances which are potentially mutagenic or carcinogenic so that public exposure to them through normal usage might represent a significant health problem. More data are needed, however, on both conventional petroleum products and synthetic fuels in this regard.

oo Pollution Control Guidance Documents - Part Of The Agency's Regulatory Strategy

Regulating new, presently non-existent energy industries, of course, presents different problems from regulating long-standing segments of United States industry. The differences are of such an extent that a unique regulatory approach is demanded. The differences arise primarily from the facts that the new energy industries are, for the most part, not yet commercialized in the United States, have potentially different effluents and emissions from those from existing pollution sources and are being developed on a telescoped time frame under a governmentally-mandated response to "the energy crisis."

Because of these circumstances, the general approach we are taking is to issue, as preregulatory multi-media guidance, a series of Pollution Control Guidance Documents, PCGDs--one for each of the major energy technologies. The focal point of each PCGD is to be a set of available control alternatives for each environmental discharge (again, for all media) along with associated performance expectations and the basis for the alternatives presented. The intent is to present guidance for plants of typical size and for each significantly different feedstock likely to be used. PCGDs will not have the legally binding authority of regulations but each will be reviewed extensively both within and outside of EPA. These documents will provide useful and realistic guidance to permit writers within EPA and the States and to the energy industry itself during its formative stages. As the energy industry develops, permits for individual installations are being issued based on best engineering judgment and, as the various PCGDs become available, permits will be prepared in light of the information the PCGDs contain. Then, as the energy industries mature and as large-scale control technology data become available, EPA will invoke its legally-binding regulatory procedures, but in a coordinated, multimedia fashion; in the water quality area, for example, this would mean the issuance of effluent guidelines and establishment of appropriate water quality standards, including consideration of related air quality and hazardous waste requirements.

oo Processes To Be Covered

Although the major objective of a PCGD is to recommend pollution control options, it will contain a great deal of background information on the energy processes themselves and on process streams and pollutant concentrations, and will, on the basis of a series of "case studies," offer specific technology-based control guidance for various kinds of energy processes. Processes to be included will cover those that are expected to be built for demonstration or commercial application first. (Table 2 shows planned process coverage for the four PCGD's currently being written). It is intended that discussion of product (e.g., low Btu coal gas) uses also will be included if use is integral with the manufacturing process. The process descriptions will detail the key features of each process and their pollution potential. If various process modifications are likely to be used, the changes in process configuration will be covered and expected changes in pollutant releases will be indicated. Pollutant releases that vary non-linearly with plant size or flow rates will also be identified and quantified to the extent possible.

The environmental control alternatives to be considered will include both end-of-pipe treatment techniques and process changes. Candidate control alternatives will be identified from existing United States and foreign bench-, pilot- and commercial-scale facilities or from different United States or foreign processes that have similar discharges. Performance and design will be included as will information on capital, operating an annualized costs. Energy usage for control alternatives will also be included. Finally, techniques for monitoring control performance will be identified. The source of all data will be clearly referenced to allow referral to original sources; uncertainties in the data will be indicated.

TABLE 2

TECHNOLOGIES FOR WHICH PCGDs ARE CURRENTLY PLANNED

Indirect Liquefaction

Lurgi Texaco Koppers Totzek (K-T)	Gasifiers
Fisher Tropsch Mobil-M Methanol	Conversion Systems

Oil Shale

- Occidental
- Rio Blanco
- Lurgi
- Paraho
- Union
- Colony

Lou Btu Gasifiers

-single bed, atmospheric, entrained gasifiers with and without sulfur control

Medium/High Btu Gasifiers

- Lurgi
- K-T
- Texaso
- Others to be decided

oo Permit Processing

Various actions have been taken which are aimed at expediting permits on energy facilities issued by EPA. We have set up our own Permits Coordination Group to carefully track permits on all energy installations, including the important synfuels ones. The Group will identify potential processing problems early and enable appropriate remedial action to be taken almost immediately. We have designated a single person in each of our Regional Offices to serve as a special point of contact for new energy facilities. These individuals have responsibility for assuring that timely review of permits for new energy facilities takes place, that industrial permit applicants are well informed as to when EPA will make decisions. Industry, especially the small and medium-sized firms, has responded very positively to this concept.

We now set target dates for permit processing based on the requirements of individual permit applications. The complexity of individual cases varies considerably and by tailoring the review schedule to each individual case, a much shorter average turn-a-round-time can be achieved than if a general schedule sufficient for all applications is used. For surface water discharge permits involved with surface mining of coal, a memorandum of understanding is being developed with the Department of the Interior's Office of Surface Mining (OSM). With this arrangement, OSM could issue a single permit under an agreement with EPA that OSM's comprehensive review procedure would also meet EPA's legislative requirements.

EPA has already issued several air pollution control permits for oil shale development. This early group of permits includes the Colony Development Operation of Exxon and TOSCO Corporations, the first commercial-scale shale retorting facility for which a permit has been granted in the United States. EPA's permit will eventually allow Colony to expand and produce 46,000 barrels per day of low sulfur distillates and other by-products. The permit will also allow Colony to construct and operate: (1) a 66,000-ton/day underground oil shale mine, (2) a surface oil shale retorting facility and (3) extensive support facilities including a 194-mile pipeline and a loading terminal. PSD permits have also been issued for the non-commercial-scale projects of Union Oil, the C-b tract (Occidental and Tenneco), and Rio Blanco (Gulf and Standard of Indiana). Another synthetic fuels facility which has received a PSD permit is the Great Plains Gasification Associates Coal gasification plant in North Dakota. This commercial facility will produce 125 million standard cubic feet per day of high Btu synthetic fuel gas.

Finally, a recent development in regulatory procedures to expedite permitting is the Consolidated Permit Program (4). The new consolidated permit regulations combine the requirements for the following five programs covered under four different Federal environmental laws:

- the National Pollutant Discharge Elimination System (NPDES) program of the Clean Water Act;
- the Prevention of Significant Deterioration (PSD) program of the Clean Air Act (but only where EPA itself is the permitting authority and only to specify permit procedures);
  Drinking Water Act (SDWA);
- the Hazardous Waste Management program under the Resource Conservation and Recovery Act (RCRA);and
- the Dredge and Fill (Section 404) program under the Clean Water Act.

The consolidated permit regulations and associated application forms provide a framework for simultaneously processing multiple EPA permit applications for the same facility. Standard information can be provided on a single form along with information required for specific permiting activities. Also, where appropriate, EPA has the ability to consolidate draft permits, public notices, public hearings and administrative records for all permitting activities for the facility or activity. These procedures should not only expedite the permitting process but also provide an opportunity for better comprehensive assessment of multimedia environmental control. The results should be more consistent and more efficient control requirements.

THE RESEARCH PROGRAM

EPA's energy and environmental research program is based on the belief that increased domestic energy production need not come at the cost of a deteriorating environment and threats to public health and welfare. The Federal Interagency Energy/Environment Research and Development Program was established to provide the information necessary to develop a scientific rationale for policies that strike a balance between ample domestic energy production, reasonable cost and environmental quality. This interagency effort is divided into two major research programs: health and environmental effects, and control technology.

The health and environmental effects program is designed to identify energy related pollutants in the environment, the mechanisms by which they move through the environment and their resulting effects on human, animal and plant populations.

The control technology program provides information on the types and quantities of pollutants released by energy supply activities and develops, or stimulates the development of, control options where necessary. A major thrust of research in the control technology program is the generation of technical and cost information on which reasonable environmental standards can be based.

EPA's research program emphasizes the generation of data necessary to support the establishment and implementation of technology-based environmental guidelines. This information will be used to assist, and ultimately minimize, environmental damage resulting from a broad array

100%

of energy fuels and processes. Those systems judged to have the greatest potential for near-tearm negative impact will receive study priority.

Over the next five years, the focus of the research program will be on the current and projected coal fuel and oil shale cycles. Over the next fifteen years, coal and oil shale production and use are expected to grow faster than any other fuel source, and they both demonstrate the potential for creating major environmental problems throughout the fuel cycle. In addition, coal is expected to be the dominant fuel employed for electricity production and will be used increasingly as a feed stock for synthetic liquids and gases.

FGD ECONOMICS IN 1980

By

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ABSTRACT

Presented in this paper is a review of recent results from EPA-sponsored flue gas desulfurization and byproduct/waste disposal economic evaluations prepared by TVA. Included are a summary of comparative capital investments and annual revenue requirements from a three-phase effort to evaluate the leading FGD processes, and similar results from three phases of sludge disposal studies. Data from a 1985 projection of FGD byproduct sulfur/sulfuric acid marketing potential are given.

A new series of FGD process evaluations is also previewed including a set of updated evaluation premises which will be utilized in the early 1980's. Examples of the effects of the revised premises on limestone scrubbing economics are shown. Finally, results are provided from a recent evaluation of limestone scrubbing in a spray tower using adipic acid, forced oxidation, and gypsum disposal by stacking.

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FGD ECONOMICS IN 1980

INTRODUCTION

Through the publication of numerous studies sponsored by EPA and other organizations, a great deal of understanding and a broadened perspective of FGD economics have been developed during the past decade. As we enter the 1980's, interest in FGD economics continues as strong as it was 10 years ago. Changes in technology, environmental regulations, economic conditions, and design philosophies all affect the projection of FGD economics to such an extent that constant reassessment is necessary. Neither the pace nor the effects of these changes can be expected to diminish soon.

The interagency EPA-TVA program to evaluate FGD economics that began in 1967 is now well into its second decade of activity. Projects to evaluate the economics of leading nonrecovery and recovery FGD processes, waste disposal processes, coal-cleaning systems, and byproduct marketing studies have all been a part of this program. Results from much of this work have been reported at earlier symposiums.

During 1980, additional results have been derived from the continuing program. This paper summarizes most of the recent published data and work in progress. First, a summary of results from three reports on comparative FGD process economics is presented. Second, a summary of information from three published reports on sludge disposal economics is given. All six of these reports utilize the same time frame (1977-1980) and design and economic premises. Reported next are the data from a 1985 projection of FGD byproduct sulfur/sulfuric acid marketing.

A new series of FGD process evaluations was begun in 1980 using an updated set of design, regulatory, and economic premises more typical of conditions to be faced in the early 1980's. Evaluation projects using a costing time frame of 1981-1984 are previewed on dry scrubbing processes, limestone process alternatives, gypsum-producing processes and ash disposal systems. The new premises are also described, as is a stepwise conversion of limestone scrubbing economics from the old premises to the new premises.

In the final portion of the paper, results are projected for an advanced limestone scrubbing process using a spray tower, adipic acid additive, forced oxidation, and gypsum stacking. This particular evaluation is for a limestone system expected to come into common usage in the future if scheduled large-scale process development is successful and environmental acceptability is proven.

Because the results presented herein are from a variety of studies using different premises, special caution should be exercised in utilizing the results. Particular attention should be paid to the different designs evaluated for the limestone scrubbing process.

FGD ECONOMIC STUDIES

In 1977 TVA began a series of FGD economic studies designed for the twofold purpose of updating previously evaluated processes and integrating evolving technologies into the EPA-TVA FGD economic studies. Three reports (1,2,3), two of which have been published, covering seven FGD systems and two processes for producing sulfur from FGD SO_2, have been prepared. The limestone and lime scrubbing processes were updated from an earlier report, as were the magnesia and Wellman-Lord scrubbing processes (4). A generic double-alkali process was included to represent this important type of nonrecovery FGD process. The citrate process and the Rockwell International aqueous carbonate process (ACP) were included as emerging sulfur-producing processes. The ACP represents two areas of new FGD technology, spray dryer FGD and the use of coal as a reducing agent to produce sulfur. The latter technology was also represented in this series of studies by the Foster-Wheeler Energy Corporation Resox® process and the Allied Chemical coal/SO_2 reduction process, both of which utilize coal to produce sulfur from SO_2. Schematic flow diagrams of all the processes evaluated in this series are shown in Figure 1.

These processes represent a range of development from established technology (the limestone and lime), through demonstration and recent commercialization (the double-alkali, citrate, magnesia, and Wellman-Lord scrubbing processes), to less-developed processes (the ACP and the Resox® and Allied coal reduction processes).

The same premises, based on a 500-MW power plant burning 3.5% sulfur coal, meeting the 1.2 lb SO_2/MBtu NSPS, and using mid-1979 capital costs and mid-1980 annual revenue requirements, were used throughout. As in other EPA-TVA economic studies, these base-case conditions were systematically varied to evaluate different fuel, power plant, and FGD conditions. In all, over 100 case variations of 9 basic FGD processes were evaluated. In addition, in recognition of the growing importance of energy in design considerations, a ground-to-ground energy evaluation was made for some of the processes.

Process Descriptions

The limestone, lime, and double-alkali processes produce a waste slurry that is disposed of in a pond. In the limestone process the flue gas is scrubbed with a slurry of ground limestone, forming calcium sulfur salts that are discarded by pumping a purge stream to a disposal pond. The lime process is similar except that a slurry of lime is used as the scrubbing medium. In the double-alkali process a solution of

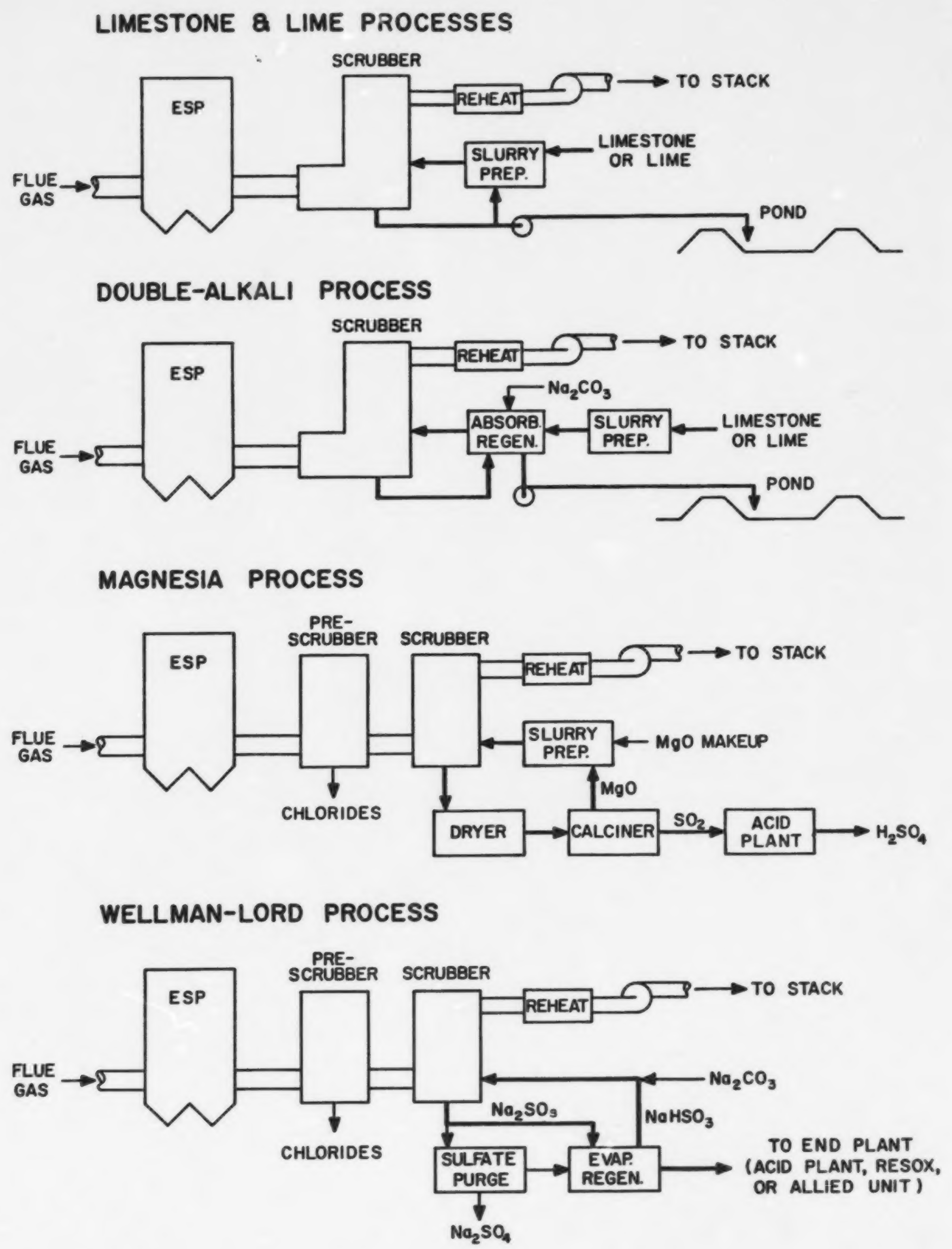
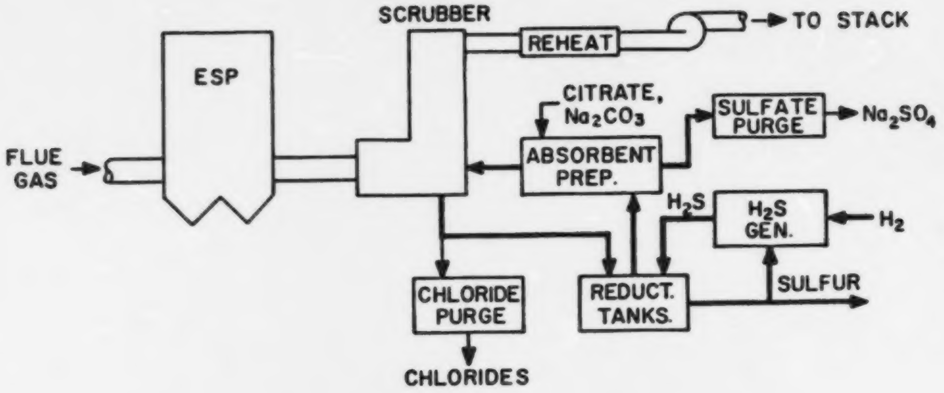
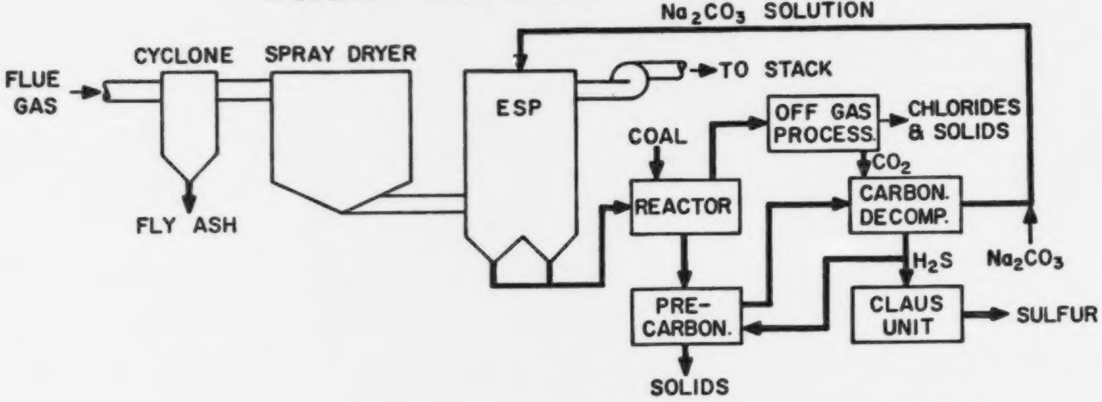

LIMESTONE & LIME PROCESSES
SCRUBBER
ESP
REHEAT
TO STACK
SLURRY PREP.
LIMESTONE OR LIME
FLUE GAS
POND
DOUBLE-ALKALI PROCESS
SCRUBBER
ESP
REHEAT
TO STACK
Na_2CO_3
ABSORB. REGEN.
SLURRY PREP.
LIMESTONE OR LIME
FLUE GAS
POND
MAGNESIA PROCESS
PRE-SCRUBBER
SCRUBBER
ESP
REHEAT
TO STACK
FLUE GAS
SLURRY PREP.
MgO MAKEUP
MgO
CHLORIDES
DRYER
CALCINER
SO_2
ACID PLANT
H_2SO_4
WELLMAN-LORD PROCESS
PRE-SCRUBBER
SCRUBBER
ESP
REHEAT
TO STACK
FLUE GAS
Na_2CO_3
Na_2SO_3
$NaHSO_3$
CHLORIDES
SULFATE PURGE
EVAP. REGEN.
TO END PLANT (ACID PLANT, RESOX, OR ALLIED UNIT)
Na_2SO_4

Figure 1. FGD process flow diagrams.

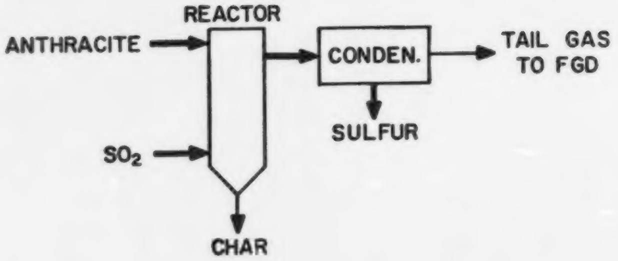
CITRATE PROCESS


SCRUBBER
REHEAT
TO STACK
ESP
CITRATE, Na_2CO_3
SULFATE PURGE
Na_2SO_4
FLUE GAS
ABSORBENT PREP.
H_2S
H_2S GEN.
H_2
SULFUR
CHLORIDE PURGE
REDUCT. TANKS.
CHLORIDES

AQUEOUS CARBONATE PROCESS


Na_2CO_3 SOLUTION
CYCLONE
SPRAY DRYER
FLUE GAS
TO STACK
ESP
OFF GAS PROCESS.
CHLORIDES & SOLIDS
COAL
CO_2
FLY ASH
REACTOR
CARBON. DECOMP.
H_2S
Na_2CO_3
PRE-CARBON.
CLAUS UNIT
SULFUR
SOLIDS

RESOX UNIT


REACTOR
ANTHRACITE
CONDEN.
TAIL GAS TO FGD
SULFUR
SO_2
CHAR

ALLIED UNIT

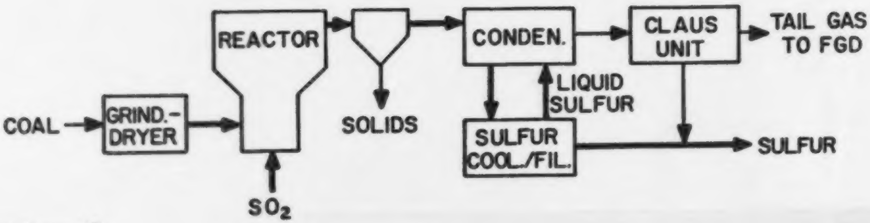

REACTOR
CONDEN.
CLAUS UNIT
TAIL GAS TO FGD
LIQUID SULFUR
COAL
GRIND.-DRYER
SOLIDS
SULFUR COOL./FIL.
SULFUR
SO_2

Figure 1 (continued)

sodium sulfite is the scrubbing medium. The spent solution is regenerated by adding lime, producing calcium sulfur salts that are discarded in a disposal pond. A slurry of magnesium oxide is used as the scrubbing medium in the magnesia process. The spent slurry is dewatered, dried, and thermally decomposed to regenerate the magnesium oxide and produce SO_2 which is converted to sulfuric acid in a conventional acid plant.

The citrate process is a wet scrubbing process using a sodium citrate solution as the absorbent. The absorbent is regenerated and the SO_x compounds reduced to elemental sulfur by liquid-phase reduction using H_2S. The H_2S is produced by reducing some of the product sulfur using natural gas.

In the Wellman-Lord process a solution of Na_2SO_3 is the scrubbing medium. Reaction with SO_x produces $NaHSO_3$ which is heated to evolve SO_2 and regenerate Na_2SO_3. Other sodium compounds, primarily Na_2SO_4, form and must be removed. Unlike the magnesia process, which produces a dilute SO_2 off-gas, the Wellman-Lord process produces an SO_2-rich off-gas more suitable for direct reduction to sulfur. In these studies it is evaluated with a sulfuric acid end plant and with the Resox® and Allied coal reduction processes.

The Resox® process consists of a vertical reactor through which rice-sized anthracite flows by gravity at a controlled rate. The SO_2-rich off-gas is mixed with controlled amounts of water and air, heated, and passed through the reactor. In a complex series of reactions some anthracite is oxidized to maintain the reaction temperature and most of the SO_2 is reduced to sulfur. A noncaking coal such as anthracite is necessary. Careful control of residence time, temperature, and SO_2:H_2O ratio is necessary to limit the thermodynamic tendency of the sulfur to go to H_2S. Sulfur is condensed from the emerging gas and the remainder is burned to convert the sulfur compounds to SO_2 and returned to the FGD system.

The Allied process uses a slightly pressurized fluidized-bed reactor containing a mixture of ground power plant coal and silica sand through which the SO_2 off-gas, mixed with a small quantity of air, is passed. Most of the SO_2 is reduced to sulfur but appreciable H_2S is also produced. The off-gas is passed through a particulate collector, a liquid sulfur scrubber to condense the sulfur, and a Claus unit to oxidize the H_2S to sulfur before the residue is incinerated and returned to the FGD system. The process also includes coal drying and grinding facilities and sulfur cooling and filtration facilities.

The ACP consists of spray dryer absorbers using a soda ash solution followed by ESP's to collect the sulfur salt particulate matter and residual fly ash not removed in upstream cyclones. The particulate matter is mixed with ground power plant coal and injected into refractory-lined reactors. Air is injected to maintain a reaction temperature of 1500°F, at which the sodium salts are molten. Most of the sulfur is reduced to the sulfide. The reactor off-gas is scrubbed to remove chlorides and ash and used as a CO_2 source. The melt overflows to a quench/dissolving tank. The dissolved melt is treated with process H_2S to form NaHS and then with process CO_2 to produce H_2S and $NaHCO_3$, which is further reacted with CO_2 off-gas to produce Na_2CO_3. The H_2S is processed to sulfur in a conventional Claus unit.

Economic Results

The base-case costs for each of the nine processes are shown in Table 1. Except for the ACP, the costs are product-related, falling into separate groupings of waste-, acid-, and sulfur-producing processes in both capital investment and first-year revenue requirements. The differences in cost between the waste-producing and acid-producing processes are essentially the costs for absorbent regeneration; ponding costs and acid plant costs do not differ greatly and raw material costs do not differ sufficiently to produce large cost differences. The higher costs for sulfur-producing processes are the result of the added costs for reduction of sulfur oxides. Here coal reduction holds a strong advantage over other fossil reducing agents. In the citrate process, 16% of the annual revenue requirements (1.06 mills/kWh of 6.44 mills/kWh) are for natural gas to produce H_2S. In contrast, reducing coal costs range from 9% (Resox®) to 4% (Allied).

TABLE 1. FGD PROCESS ECONOMIC COMPARISONS

	Mid-1979 capital investment, $/kW	Mid-1980 first-year revenue requirement, mills/kWh
Waste-Producing Processes		
Limestone	98	4.02
Lime	90	4.25
Double alkali	101	4.19
Sulfuric Acid Processes		
Magnesia	132	5.08
Wellman-Lord/sulfuric acid	131	5.11
Sulfur Processes		
ACP	119	4.81
Wellman-Lord/Resox	138	6.03
Wellman-Lord/Allied	141	5.94
Citrate	143	6.44

The anomalous capital investment of the ACP results from a credit for the unnecessary separate fly ash ESP's and from the intrinsic chloride purge from the reducer off-gas quench. If no ESP credit is given (as in an existing plant with ESP's in place) its capital investment becomes 137 $/kW. Similarly, if no chloride removal is necessary in the wet processes, these process costs are reduced about 10 $/kW. Under these conditions, the ACP becomes the highest in capital investment. Specific power plant conditions are thus important in the comparative capital investments of the regeneration processes. In first-year revenue requirements the lower costs for the ACP are less site specific. It has low raw material costs and low utility costs that prevail regardless of specific fuel and power plant conditions.

Ground-to-Ground Energy Assessment

As a part of this series of FGD studies, a ground-to-ground energy assessment of the limestone, lime, and magnesia processes was made. This consisted not only of the FGD energy requirements but the energy consumed in mining, processing, and transportation of the raw materials, the disposal of wastes, and an energy credit for the sulfuric acid produced. The assessment represents, in a sense, the energy removed from a hypothetical energy reservoir because of the operation of the FGD systems. A credit is given for the sulfuric acid because it replaces acid that would be produced from sulfur, and thus the energy that would have been consumed in mining and transporting the sulfur and producing the acid. The results are shown in Table 2 and Figure 2.

TABLE 2. FGD GROUND-TO-GROUND ENERGY REQUIREMENTS

Function	Btu/lb sulfur removed		
	Limestone	Lime	Magnesia
Mining	438	356	25
Absorbent processing	-	6,198	161
Transportation	176	143	33
FGD	14,042	13,165	26,387
Waste disposal	22	15	-
Total	14,678	19,877	26,658
Byproduct credit	-	-	(5,491)
Net total	14,678	19,877	21,115
Btu/kWh	291	395	420
% gross power unit output	3.2	4.4	4.7

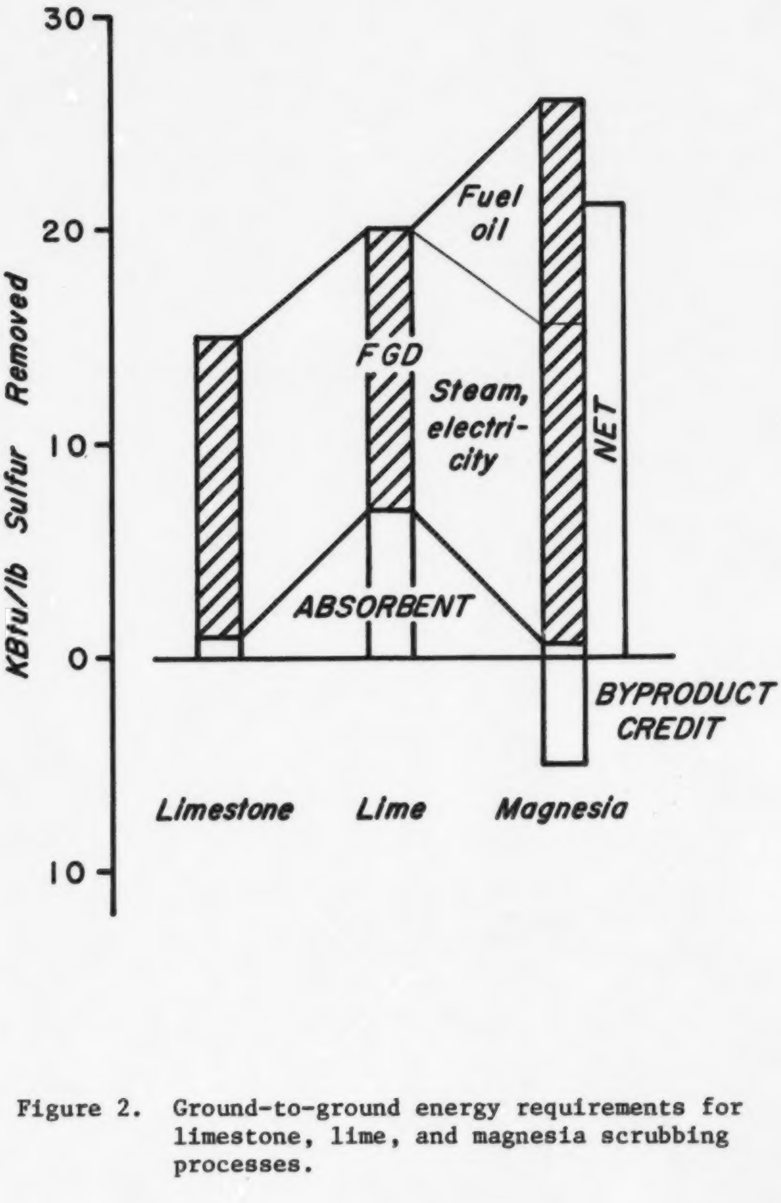


Figure 2. Ground-to-ground energy requirements for limestone, lime, and magnesia scrubbing processes.

The ground-to-ground energy comparison shows considerably different relationships than comparison of FGD energy requirements alone. The FGD energy requirements of the magnesia process (typical of regeneration processes) are about twice those of the limestone and lime processes. The absorbent energy requirements are low for the magnesia process because only makeup magnesia is used. In contrast, the lime process, which has the lowest FGD energy requirements, has much higher energy requirements when the energy for calcining lime is included. With the byproduct credit included, the magnesia process is not appreciably more energy intensive than the lime process.

Energy requirements cannot, of course, be directly related to FGD costs. Energy consumed in absorbent production and transportation, for example, is seen only indirectly, as it affects raw material costs. In addition, the form of the energy may have an important effect on costs. The magnesia process uses fuel oil for over one-third of its energy requirements whereas almost all of the limestone and lime energy requirements are met with coal. The significance of these differences on costs is dicussed further in the byproduct marketing portion of this paper.

FGD WASTE DISPOSAL ECONOMICS

Also during the past three years, TVA has conducted a series of evaluations for EPA on the economics of disposal processes for flue gas cleaning wastes. The first three studies (5,6,7) deal with the disposal of fly ash and scrubber wastes from limestone/lime FGD systems. In all, seven disposal methods were evaluated covering a range of existing or potential disposal options of the late 1970's. All of the evaluations were based on the same premises, using as the basis a 500-MW power plant burning a 3.5% sulfur eastern coal and scrubbing with a limestone slurry to meet the then-existing 1.2 lb S/MBtu NSPS. In addition, over 175 case variations representing various power plant, fuel, waste treatment, transportation, and disposal site conditions were evaluated. Schematic flow diagrams of the processes are shown in Figure 3.

Except for the gypsum process, the scrubber waste consists of a 15% solids slurry with a sulfur species composition of 85% $CaSO_3 \cdot 1/2H_2O$ and 15% $CaSO_4 \cdot 2H_2O$. Fly ash is included in the slurry except for the sludge - fly ash blending and Dravo landfill processes. In dewatering, 30% solids from the thickener and 60% solids from the filter is used. For the gypsum process essentially all the sulfur is $CaSO_4 \cdot 2H_2O$ and the filtered solids is 80%.

Process Descriptions

The untreated ponding case assumes that the effluent is pumped directly to an earthen-diked pond. The Dravo, IUCS, and Chemfix processes are all commercial fixation processes using somewhat different approaches to treat dewatered FGD sludge. All depend on additives that produce

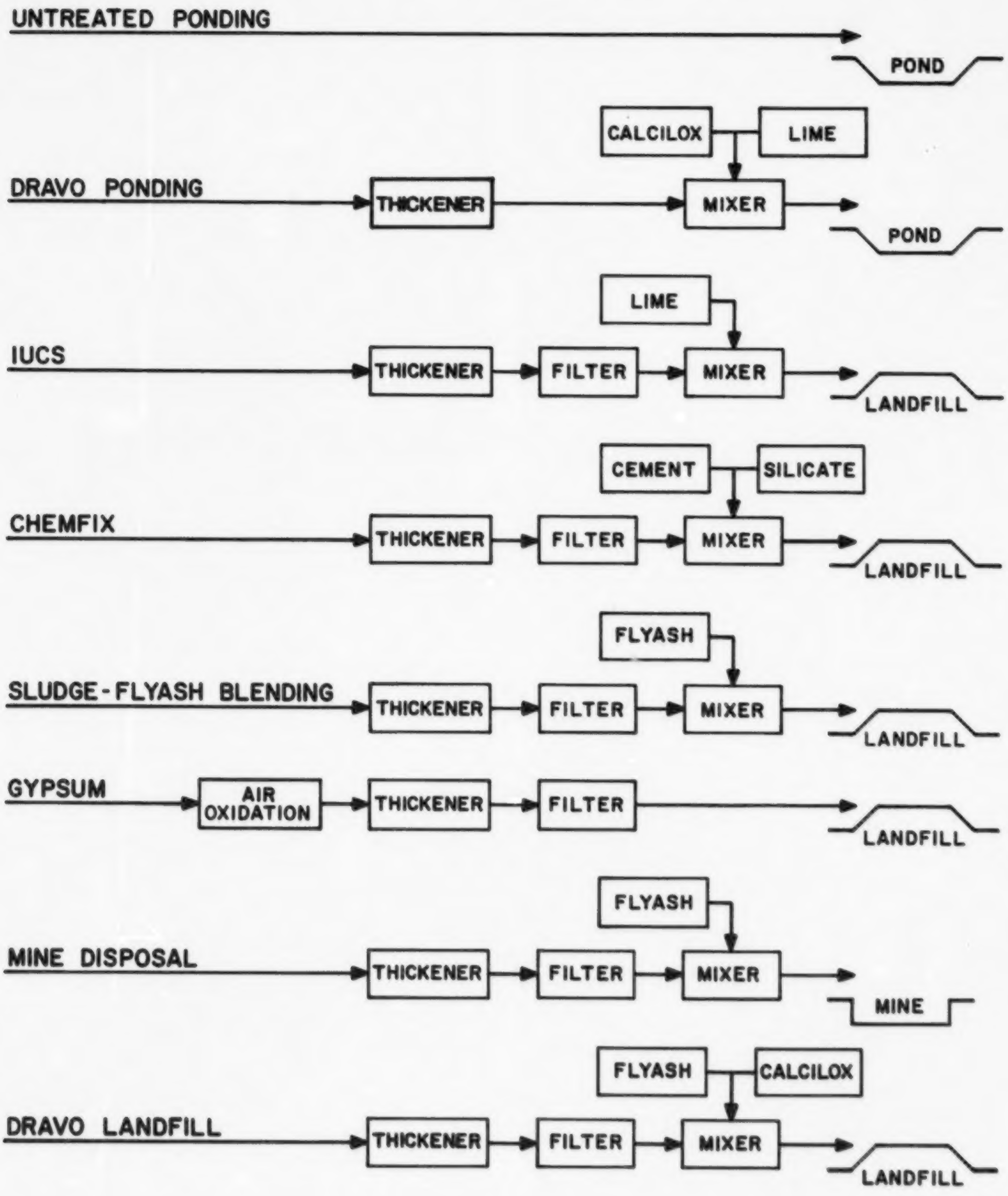

UNTREATED PONDING
POND
CALCILOX
LIME
DRAVO PONDING
THICKENER
MIXER
POND
LIME
IUCS
THICKENER
FILTER
MIXER
LANDFILL
CEMENT
SILICATE
CHEMFIX
THICKENER
FILTER
MIXER
LANDFILL
FLYASH
SLUDGE-FLYASH BLENDING
THICKENER
FILTER
MIXER
LANDFILL
GYPSUM
AIR OXIDATION
THICKENER
FILTER
LANDFILL
FLYASH
MINE DISPOSAL
THICKENER
FILTER
MIXER
MINE
FLYASH
CALCILOX
DRAVO LANDFILL
THICKENER
FILTER
MIXER
LANDFILL

Figure 3. Process flow diagrams.

cementitious chemical reactions. The types and quantities of the additives and the degree of dewatering can be controlled to produce a soillike material over a curing period of hours or months. The Dravo process uses their product Calcilox,® a processed blast furnace slag, sometimes with lime or fly ash, or both. Depending on the degree of sludge dewatering and materials added, the treated material is pumped to permanent or temporary pond storage or it is hauled to disposal after a curing period. The IUCS process uses lime and fly ash blended with dewatered sludge to produce a soillike solid. The Chemfix process uses portland cement and sodium silicate blended with dewatered sludge to produce a soillike solid. The process is said to provide an encapsulation that reduces leaching. For comparison, a sludge - fly ash blending process without purchased additives, is included. The gypsum process differs in that air oxidation equipment is added to the scrubber loop, permitting production of the more easily dewatered and denser $CaSO_4 \cdot 2H_2O$. It is assumed this material can be dewatered and handled as a solid without stabilization or fixation with additives. Finally, a process using the sludge - fly ash blending process with disposal in a surface mine is evaluated.

Economic Results

Cost breakdowns of the base cases by processing areas were made, as shown in Table 3, to facilitate identification of cost elements and comparison of different disposal processes. The sludge - fly ash blending process, the mine disposal process, and the Dravo landfill process require inclusion of ESP costs for comparison with the other processes.

In those cases in which fly ash is collected separately the cost of ESP units and their operation is a major component of the waste disposal costs. In comparison, simultaneous fly ash removal results in relatively modest increases in thickening and filtration costs. Separate collection of fly ash is, of course, possible with all of the processes evaluated and would require similar costs for all processes. In comparison of landfill disposal practices having separate fly ash collection, cost differences would largely be reduced to the raw material portion of the cost breakdown.

For the processes using purchased fixatives, raw materials are an important element of both capital investment and first-year revenue requirements. Fly ash handling is also a relatively expensive element. The advantage of a single fixative is illustrated by comparison of raw material costs for processes that use two additives with processes that use one. Thickening is the largest capital investment cost element, excluding ESP costs, for all of the nonponding processes. It is also a large cost element in annual revenue requirements. Filtration is also a large cost element, though considerably less so than thickening. Dewatering costs for the gypsum process are lower than the other simultaneous fly ash - FGD waste filtration processes because of the predicted superior filtration characteristics of the high-sulfate sludge. Mixing costs are a minor part of both capital investment and annual revenue requirements.

TABLE 3. MODULAR COSTS BY PROCESSING AREA FOR EIGHT DISPOSAL PROCESSES

	Other	Raw materials	Thickening	Filtration	Mixing	Storage	Disposal	Total	
	Capital investment by processing area, $/kW								
Ponding					1.4		33.0	34.4	
Dravo ponding		9.0	8.4		0.5		30.3	48.2	
IUCS		4.2	8.5	4.1	1.1		3.5	21.4	
Chemfix		8.5	9.1	4.8	1.6		3.1	27.1	
Sludge-fly ash blending	19.2[a]	4.4	6.3	2.5	0.9		3.1	36.4	
Gypsum	4.6[b]		5.2	3.0			2.6	15.4	
Mine disposal	19.2[a]	4.4	6.2	2.5	0.9		2.0	35.3	
Dravo landfill	19.2[a]	6.2	6.0	2.2	0.8	1.1	3.8	39.4	
	First-year revenue requirements by processing area, mills/kWh								$/ton dry waste
Ponding					0.14		0.80	0.94	8.1
Dravo ponding		0.91	0.24		0.03		0.74	1.91	15.3
IUCS		0.44	0.29	0.18	0.06		0.54	1.51	12.6
Chemfix		0.97	0.29	0.19	0.06		0.49	2.00	15.9
Sludge-fly ash blending	0.56[c]	0.22	0.24	0.11	0.05		0.45	1.64	9.3
Gypsum	0.29[d]		0.29	0.16			0.44	1.18	7.9
Mine disposal	0.56[c]	0.22	0.25	0.11	0.05		0.36	1.54	8.2
Dravo landfill	0.56[c]	0.57	0.22	0.10	0.05	0.03	0.47	2.00	11.9

Basis: 500-MW power plant, 127,500-hour life, 7,000 hr/yr revenue requirement basis; 3.5% S, 16% ash coal; fly ash removal in scrubber where cost is not shown. Limestone scrubber, 1.5 stoichiometry, 15% solids waste to disposal system.

[a] $9,614,000 ESP cost for separate fly ash collection.

[b] $2,303,000 air-oxidation modifications.

[c] $1,975,000 ESP operating costs.

[d] $1,005,000 air-oxidation operating costs.

Transportation and disposal site costs illustrate fundamental differences between ponding and landfill disposal methods. Capital investment for ponding transportation and disposal site costs is an order of magnitude greater than the capital investment for landfill transportation and disposal site operations. Capital investment for transport lines is also an important element in ponding. Among the landfill and mine disposal processes, transportation and disposal site costs are a relatively minor element of total capital investment.

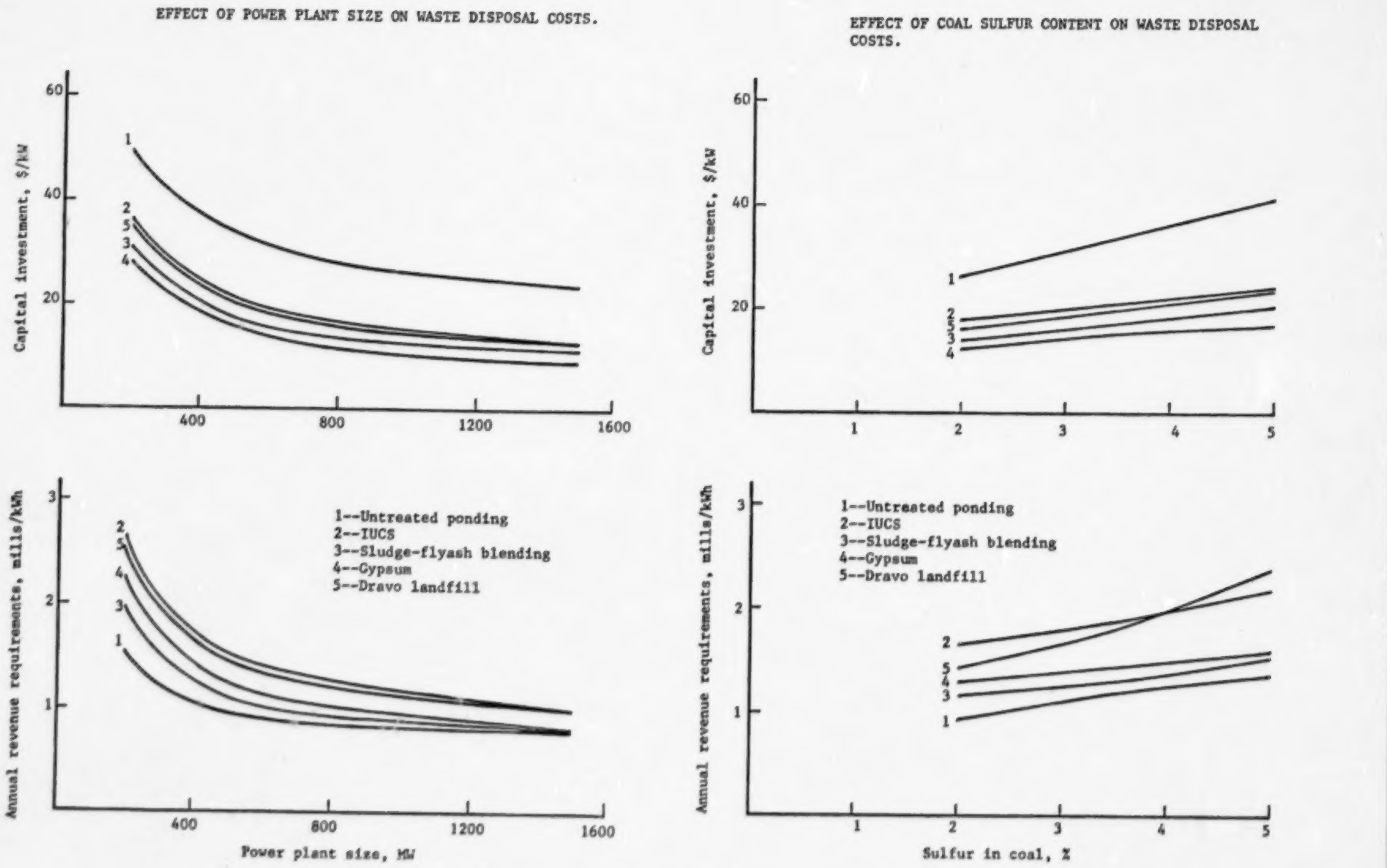
First-year revenue requirements for ponding transportation and disposal site costs are also higher than those for landfill and mine disposal although the differences are less pronounced. About two-thirds of the annual revenue requirement direct costs for ponding transportation and disposal site operations consist of pond maintenance. Transportation of the waste is a relatively minor cost element. In contrast, about four-fifths of the annual revenue requirements direct cost for landfill and mine disposal transportation and disposal site operations is for labor and supervision, much of it for loading and hauling.

In overall comparison of the processes evaluated, the most important capital investment cost elements are separate fly ash collection, raw material handling, thickening, and pond construction. Large cost elements in first-year revenue requirements are separate fly ash collection, raw material purchase and handling, and disposal.

The most important variations from the base-case conditions affecting costs are power plant size, coal sulfur and ash content, and transportation distance to the disposal site, as shown in Figure 4. Coal sulfur content affects costs both through the volume of waste to be processed and disposed of and, for processes using fixatives, the quantity of fixative required. Costs for the disposal processes increase at different rates with increasing sulfur content, depending on the relative influence of these factors. Fixation processes increase in cost more rapidly than the processes that do not use purchased fixatives. Distance to the disposal site illustrates an important difference between the ponding and landfill processes. Ponding investment costs increase dramatically as the distance increases to 5 and 10 miles. In contrast, transportation costs for landfill processes decrease more slowly with distance. The relatively small cost advantages of mine disposal are lost in higher transportation costs if the comparison is made between a landfill onsite and a mine over a few miles from the power plant. From a purely economic viewpoint, mine disposal requires very close proximity of power plant and mine for its economic advantages to be realized.

BYPRODUCT MARKETING

The EPA-sponsored FGD byproduct marketing system began as a limited production-marketing model for sulfuric acid in the early 1970's (8). Several expansions of the methodology led in 1978 to the basis of the present system (9), a comprehensive analysis of the potential of FGD



EFFECT OF POWER PLANT SIZE ON WASTE DISPOSAL COSTS.

EFFECT OF COAL SULFUR CONTENT ON WASTE DISPOSAL COSTS.

Figure 4. Effect of process variations on disposal costs.

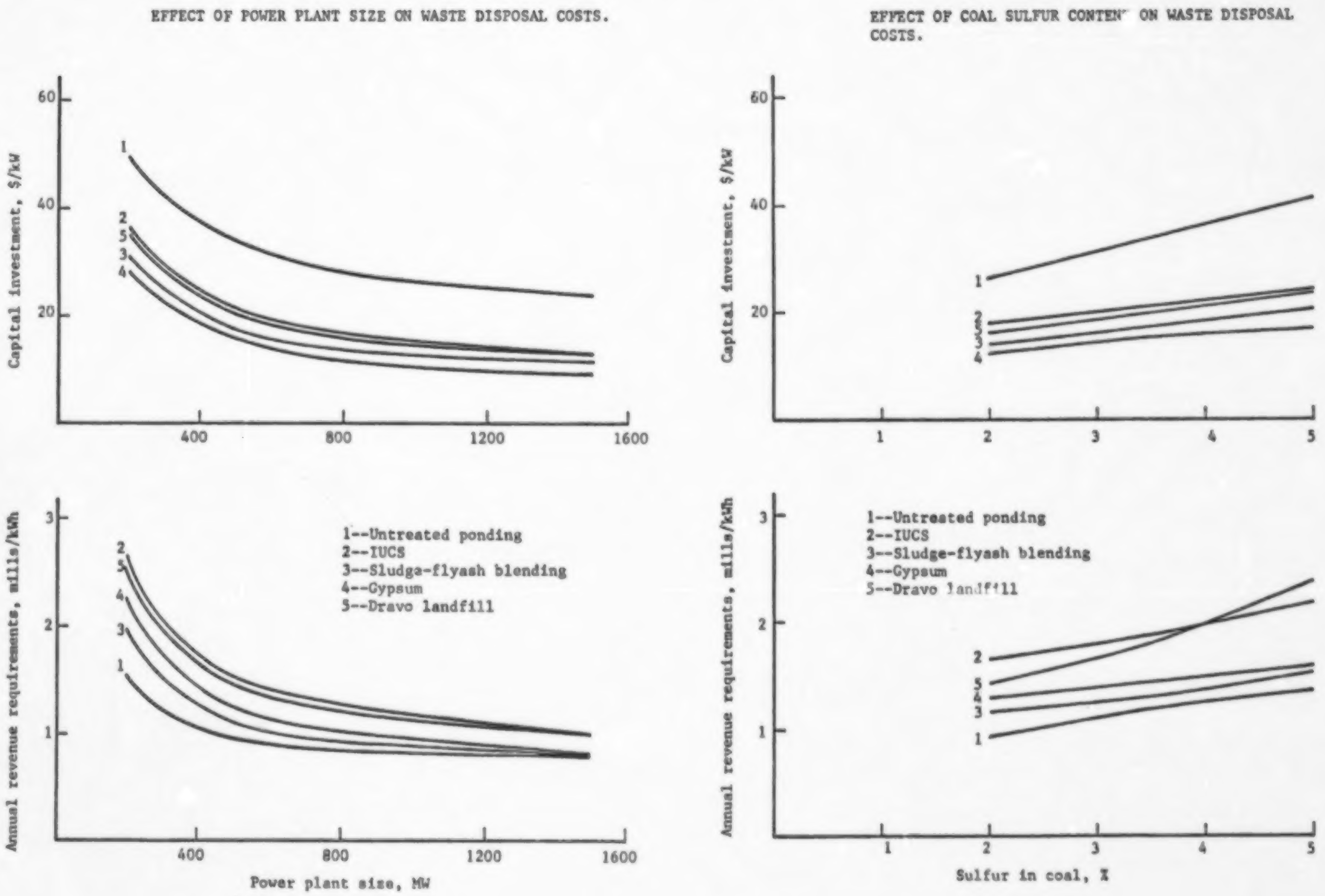
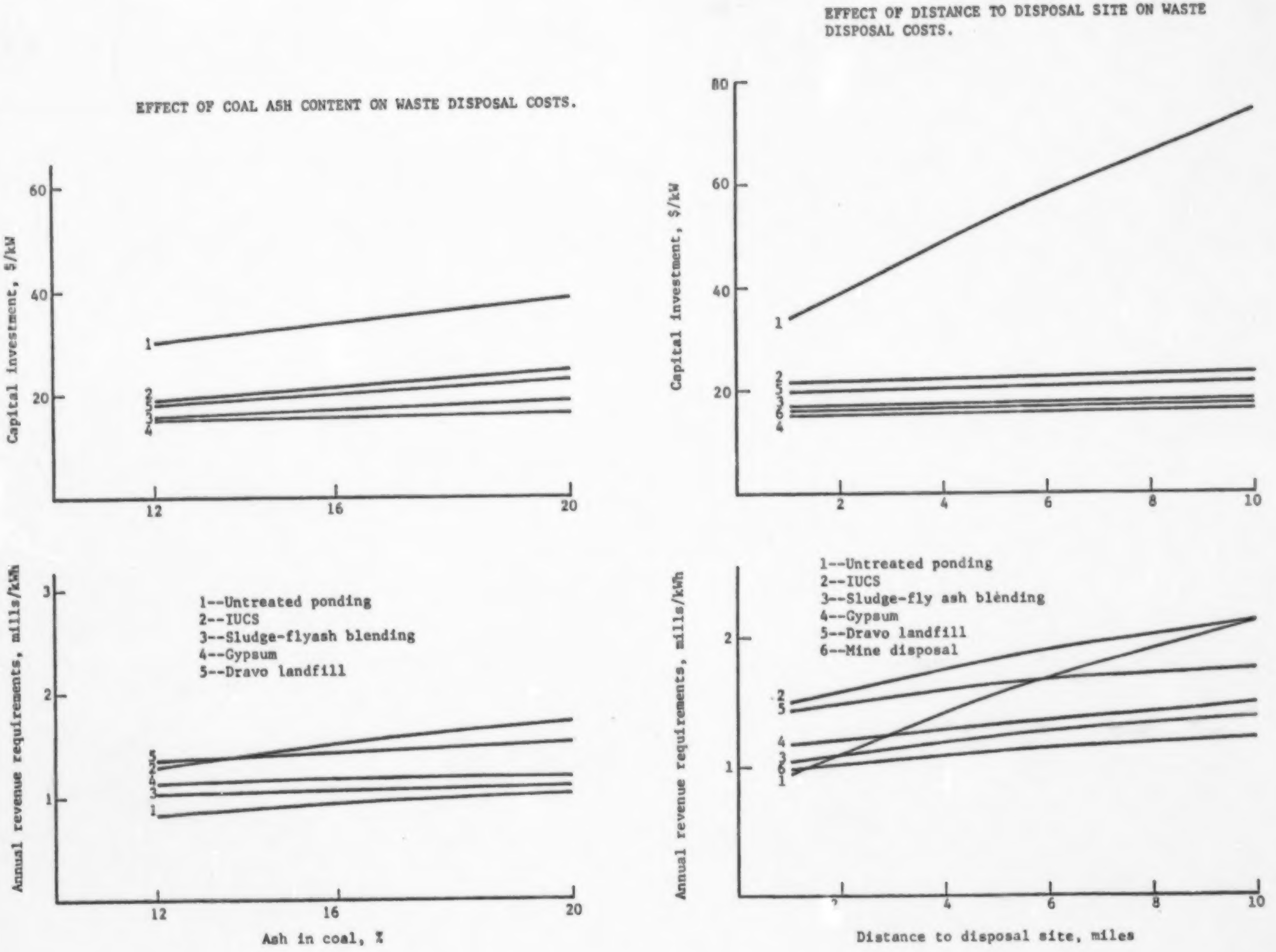


Figure 4. Effect of process variations on disposal costs.



EFFECT OF COAL ASH CONTENT ON WASTE DISPOSAL COSTS.

EFFECT OF DISTANCE TO DISPOSAL SITE ON WASTE DISPOSAL COSTS.

Figure 4 (continued)

byproduct sulfur and sulfuric acid production and marketing by U.S. electric utilities. Basically the system compares low-sulfur fuel and regeneration and waste-producing FGD costs for existing and planned U.S. utility power plants, determines FGD byproduct revenue from sales to U.S. sulfuric acid plants, and determines the mix of strategies that results in the least-cost option and the highest total revenue from FGD byproduct sales. FERC and published utility data, transportation data, and U.S. sulfuric acid plant data are used. TVA process economics, scaled to projected power plant operating conditions, determine FGD costs.

An updated projection of FGD sulfuric acid marketing potential for 1983 was published in 1979 (10), as was a users manual for the computerized system (11). The 1983 projection also contained a manually prepared forecast of FGD sulfur marketing potential. Several trends became apparent in the 1983 projection: rapidly evolving FGD technology; disproportionate fuel cost changes, particularly for petroleum products; changes in historical patterns of utility coal use and sulfur and sulfuric acid production; and evolving environmental legislation promised to influence earlier patterns of FGD byproduct production.

Developments in FGD, such as the recognition that chloride control was necessary in some cases for regeneration processes to prevent loss of absorbent effectiveness, special purge systems, and severe corrosion problems, altered FGD costs. New technologies, such as spray dryer FGD and coal reduction, promised further changes. The type of fuel used in the FGD process was also becoming an important economic factor. The growing importance of secondary sulfur and sulfuric acid production was seen to be a potentially important consideration. Legislation such as RCRA and the 1979 NSPS revisions, restricting waste disposal options and the use of low-sulfur fuel, would be important in FGD economics in the 1980's. It was also apparent that the usefulness of these projects would be increased by extending them further into the future, on a scale similar to the time period required for power plant planning and construction.

Beginning in late 1979, a projection for 1985 was started. Although a 1990 projection would have been more desirable, availability of data, particularly on power plant construction, coal use, and fuel costs, precluded a projection beyond 1985 at that time. Numerous system changes were made, including updated FGD technologies (limestone throwaway, magnesia to acid, and ACP for sulfur), a general updating of power plant, transportation, and acid plant data, inclusion of a spray dryer FGD sulfur-producing process, and inclusion of Canadian sour gas sulfur as a market factor in the upper United States. The results, which were published this year (12), showed a number of changes from previous projections.

The combined sulfur and sulfuric acid market for 1985 was projected to be 165,000 tons of sulfur from 11 power plants and 554,000 tons of sulfuric acid from 6 power plants. The total benefits for the electric utility and sulfuric acid industry were about $20,000,000. The results, shown in Table 4, differ considerably from the 1983 projection, which showed 1,200,000 tons of sulfuric acid but no sulfur.

Several factors are important in both the total FGD byproduct production projected and the sulfur-sulfuric acid mix. Most of the production of both comes from new plants projected for a 1985 startup, which were assigned to regulation under the 1979 revised NSPS for modeling purposes. In addition, inclusion of fixation and landfill disposal in the limestone scrubbing process used for the waste-producing FGD option enhances the FGD byproduct option, although limestone scrubbing remains the predominate FGD option.

Sulfuric acid production was reduced by several factors, among which increased costs for the magnesia process used in the FGD model were most important. Inclusion of provisions for chloride control and the cost of fuel oil in the process were particularly important. The increase in potential FGD sulfur production stems largely from the use of a spray dryer recovery FGD process based on the Rockwell International aqueous carbonate process. Reduced costs in the form of simultaneous fly ash and particulate sulfur salts collection and the use of coal as the reducing agent, were important factors. In maximizing the combined sulfur-sulfuric acid market, all of which is assumed to be sold to sulfuric acid plants, alternate markets for sulfur were also more prevalent than those for sulfuric acid.

The 1985 projection indicates several factors that will have important influences on FGD byproduct production by the late 1980's. Environmental legislation affecting waste disposal practices and the restricting use of low-sulfur coal as a compliance strategy could enhance the economic attractiveness of regeneration FGD processes. The economics of byproduct FGD processes that use coal as the fuel in the regeneration-manufacturing process will be more favorable than those using oil or natural gas. Similarly, processes that combine flue gas cleaning functions, such as fly ash and sulfur salt collection, will have important economic advantages.

Fuel Oil Price Escalation

An interesting aspect of FGD economics in the past few years, as the cost basis has been projected into the 1980's, is the disproportionate effect of energy costs. This is particularly apparent in the byproduct marketing studies, which are projected further into the future than most FGD economic studies. In the 1985 projection a 15% annual inflation rate for No. 6 fuel oil was used, based on petroleum cost projections available in early 1980. As an illustration of the effect of this rate on costs, equivalent cost increases for fuel oil, natural gas, and coal are shown below.

TABLE 4. 1985 PROJECTION OF THE PRODUCTION AND DISTRIBUTION OF FGD SULFUR AND SULFURIC ACID

Power plant location	Tons	Consumer location	Tons
Sulfur			
Staten Island County, NY	7,000	Newark, NJ	7,000
Martin County, FL	28,000	Pierce, FL	28,000
Washington County, FL	20,000	Dothan, AL	7,000
		White Springs, FL	13,000
Sherburne County, MN	8,000	Dubuque, IA	8,000
Westmoreland County, PA	24,000	North Bend, OH	8,000
		Copley, OH	16,000
Montgomery County, MD	10,000	Baltimore, MD	10,000
Shelby County, AL	12,000	Tuscaloosa, AL	12,000
Williamson County, IL	11,000	East St. Louis, IL	11,000
Rusk County, TX	9,000	Fort Worth, TX	9,000
Henderson County, TX	7,000	Fort Worth, TX	7,000
Armstrong County, PA	29,000	Cleveland, OH	29,000
	165,000[a]		165,000[a]
Sulfuric Acid			
Person County, NC	103,000	Richmond, VA	36,000
		Wilmington, NC	26,000
		Norfolk, VA	41,000
Jasper County, IL	122,000	Tuscola, IL	122,000
Pike County, IN	51,000	Indianapolis, IN	51,000
Northhampton County, PA	182,000	Deepwater, NJ	95,000
		Edison, NJ	74,000
		Gibbstown, NJ	13,000
Delaware County, PA	53,000	Gibbstown, NJ	53,000
Titus County, TX	43,000	Shreveport, LA	43,000
	554,000[b]		554,000[b]

a. The potential revenue/savings to both industries combined is projected to be as much as $10,000,000 for an approximate average of $60/short ton of sulfur.

b. The potential revenue/savings to both industries combined is projected to be as much as $10,500,000 for an approximate average of $19/short ton of sulfuric acid.

Annual price escalation, %	Equivalent price increase, 1979-1985 No. 6 fuel oil, $/gal	Natural gas, $/kft^3	Coal, $/ton
5	0.20	1.37	30.13
15	0.79	5.29	116.32
25	1.69	11.33	249.36

To equal the price increase projected for fuel oil, for example, the price of coal would have to increase over 100 $/ton. Processes such as the magnesia process that use fuel oil are thus placed at a disadvantage compared with processes such as the ACP using coal.

The effect of fuel oil price escalation on the cost of FGD sulfuric acid is shown in Figure 5. The effect is twofold, first in FGD costs and second in the avoidable production costs to acid producers. This is a cost calculated by the byproduct marketing system to determine the price of FGD acid at each acid plant. It represents the break-even point between buying FGD acid to meet marketing requirements and producing acid. In shutting down an acid plant, however, steam production is lost and normally must be replaced by a boiler. Because of size, this logically would be an oil-fired boiler. High fuel oil price escalation rates thus decrease avoidable production costs, resulting in the need of a higher acid price margin to make the purchase of FGD acid economical.

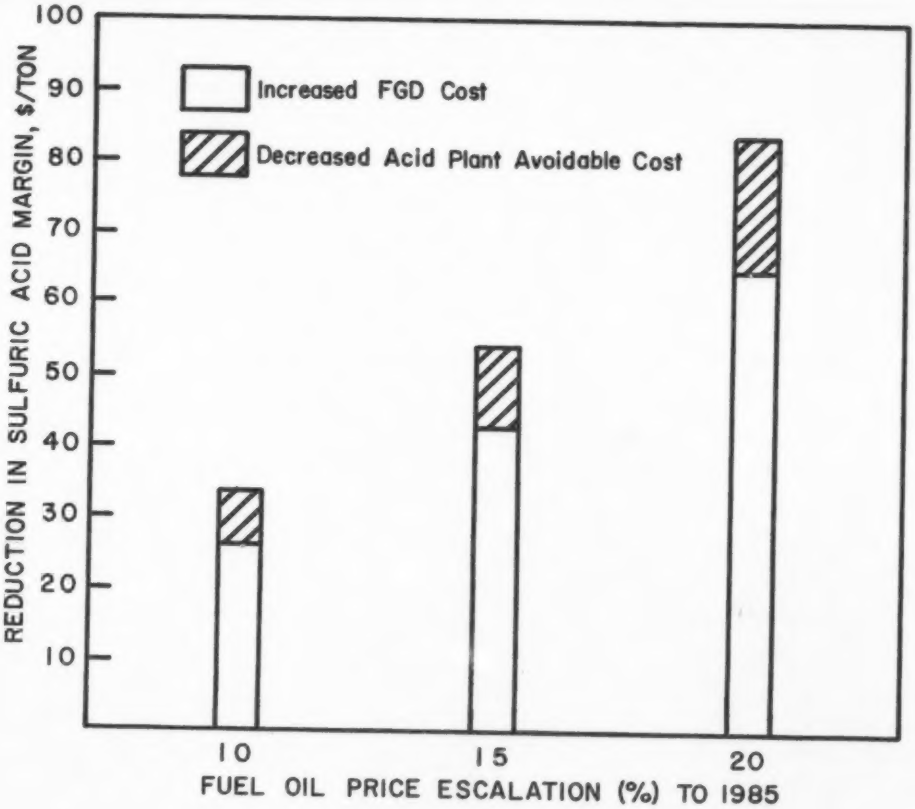


Figure 5. Reduction in potential FGD sulfuric acid margin with No. 6 fuel oil annual price escalation.

FGD AND SOLID WASTE PROCESS EVALUATIONS IN PROGRESS

With the completion of the 1977-1980 series of SO_x control and FGD solid waste process designs and evaluations, plans were made for extension of the series to other important FGD and waste disposal processes not yet evaluated. During the planning cycle, dry scrubbing processes were just beginning to capture strong interest. Therefore, the first new study for the 1980's was a preliminary economic evaluation of this technology. The first report on a lime spray dryer system for a western low-sulfur coal application was published during early 1980 (13). A second more detailed report summarizing current dry FGD process technology and the economics for both low- and high-sulfur coal will be published soon (14). T. A. Burnett will present results from these reports in a paper to be presented later in the symposium.

A second project is now underway to prepare a report summarizing the designs and economics of wet limestone-lime processes which have been studied at the EPA-TVA Shawnee Test Facility. Thirteen different process variations included in this report are listed below.

1. Turbulent Contact Absorber® (TCA) - Onsite ponding
2. TCA - Forced oxidation - Landfill
3. TCA - Forced oxidation Adipic acid - Landfill
4. TCA - Forced oxidation - MgO - Landfill
5. Spray Tower (ST) - Onsite ponding
6. ST - Forced oxidation - Landfill
7. ST - Forced oxidation - Adipic acid - Landfill
8. ST - Forced oxidation - MgO - Landfill
9. Venturi-Spray Tower (V-ST) - Onsite ponding
10. V-ST - Forced oxidation - Landfill
11. V-ST - Forced oxidation - Adipic acid - Landfill
12. V-ST - Forced oxidation - MgO - Landfill
13. Venturi - Forced oxidation - Adipic acid - Landfill

The final report should be available during 1981.

A third project, which is about half completed, is a study of three leading gypsum-producing FGD systems. The Dowa process, which was developed in Japan on oil-fired boilers, is being marketed in the United States by UOP and has been tested on a 10-MW prototype at Shawnee, is one of the processes. The Saarberg-Hölter process, a German-developed system marketed by Davy-McKee in the United States, is the second process. The third system is a limestone spray tower using adipic acid addition, forced oxidation, and gypsum stacking for waste disposal. The report for this project is expected to be ready for distribution in mid-1981. There are other gypsum-producing processes being developed for commercial use; it is hoped that these can be evaluated in a future study.

The last defined project now underway in the expanded series is an evaluation of ash disposal systems and practices for coal-fired power plants. The draft report for this project has been prepared and publication is expected shortly.

The ash disposal methods evaluated in this study are represented by five base-case processes based on major utility ash disposal practices. Four base cases represent disposal of noncementitious eastern coal ash. They consist of (1) direct sluicing of combined fly ash and bottom ash to separate ponds with once-through (nonrecycled) water, (2) the same system with recycled transportation water, (3) direct sluicing of fly ash and bottom ash to temporary ponds, followed by excavation and trucking of both to a common landfill, and (4) collection of bottom ash in dewatering bins from which it is trucked to a separate landfill and collection of fly ash in dry storage silos from which it is trucked to a separate landfill.

The fifth base case represents a situation in which the power plant is burning a western-type coal which contains appreciable calcium, making the ash subject to spontaneous cementitious reactions that affect handling properties. The handling and disposal system is designed to forestall these reactions by keeping the ash dry until shortly before placement at the disposal site.

NEW PREMISES

The FGD and waste disposal studies that are now in progress are based on new design and economic premises. During the 1977-1980 series of studies it was recognized that changing economic conditions, fuel use patterns, developments in economic evaluation techniques, and, particularly, developments in FGD technology and environmental legislation justified revision of the TVA design and economic premises. Consequently, TVA began studies that led to the adoption of new economic premises in 1979. During this period numerous discussions were held with EPA, EPRI, and with other TVA organizations concerned with the use of these premises.

Design Premises

Essentially the same power plant conditions are retained. For the base case these are a new, midwestern, 500-MW, pulverized-coal-fired, dry-bottom boiler. The heat rate is increased from 9,000 to 9,500 Btu/kWh and the excess air is increased from 33% to 39%, however. The sulfur content of the coal remains at 3.5% but the heating value is increased from 10,500 to 11,700 Btu/lb. The operating schedule is also changed to 5,500 hr/yr for 30 years. A constant annual operating time is used to facilitate levelizing of lifetime costs.

Major changes were made in the FGD design premises to reflect current regulations and to improve process reliabilities. Required SO_2 removal efficiency is now based on the 1979 NSPS. For the base-case coal these require an 89% removal efficiency instead of the 79% needed to meet the 1971 NSPS used in the old premises. In keeping with current design trends a spare absorber train and provisions for emergency bypass of 50% of the total flue gas are included. The old premises contained no spare absorber or bypass provisions. In addition, ID booster fans, instead of FD booster fans, are used in the new designs. For nonrecovery processes both pond and landfill waste disposal methods are revised to reflect more recent environmental concerns. These are primarily based on RCRA Subtitle D (nonhazardous waste) guidelines and include provisions for such factors as seepage and runoff control, security, monitoring, and reclamation.

FGD process design features are usually based on technology prevailing at the time of the study. The limestone scrubbing process is, however, somewhat of a premise adjunct since it is used so frequently as a basis of comparison in FGD studies. This process serves as an example of the changes in FGD technology that have occurred over the past few years. The current limestone process differs from the old process used in the 1977-1980 studies in several respects. A spray tower instead of a mobile-bed absorber, forced oxidation to gypsum, and landfill waste disposal are now included in the basic system. The use of a spray tower results in a lower gas velocity of 10 ft/sec instead of the 12.5 ft/sec used in the old process with a mobile-bed absorber.

The new limestone scrubbing process represents several industry trends in limestone scrubbing that have become evident in recent years. The use of a spray tower instead of more complicated mobile-bed and venturi - spray scrubbers has become common. The simpler spray tower is expected to provide greater reliability and require less maintenance although these have not been quantified in practice. The problem of waste disposal has also been addressed, both by increasing use of stabilization, fixation, and landfill disposal techniques and by other methods of producing a more tractable waste, such as oxidation to gypsum.

The use of a spray tower, air oxidation, and landfill disposal in the new process recognizes these trends. The process is based in part on continuing test work on spray towers, forced oxidation, and waste dewatering at the EPA-sponsored test facility at the Shawnee Steam Plant. Like the previous limestone scrubbing process, however, it is generic and incorporates general industry information as well as data from Shawnee.

Economic Premises

Numerous changes were also made in the economic premises. Specific provisions for sales tax, freight, and overtime for construction delays are included. The method of calculating indirect capital investment is simplified and modified to more accurately reflect complexity of engineering and construction costs of processes evaluated. Contingencies and allowances for modification after startup are also defined as process-specific variables reflecting degree of development and established technology. Provision for recognition of anticipated royalties is also made. Land prices and interest during construction are increased.

First-year revenue requirements are now calculated using levelized capital charges (30-year life, capital recovery factor, 6% per year inflation and 10% per year cost of money, discounted to the first year) instead of the average capital charges used in the old premises. In addition, levelized lifetime revenue requirements are also calculated to represent inflated and discounted costs over the life of the system.

The base years for capital investment and first-year revenue requirements are also advanced to 1982 and 1984 respectively. A project construction period from 1981 to 1983 is now assumed, with plant startup in early 1984.

COST COMPARISON OF OLD AND NEW PREMISES

The key old and new design and economic premises for evaluation of the limestone scrubbing process are shown in Table 5. A stepwise cost transition from the old premises and technology to those for the new limestone scrubbing evaluation is shown in Table 6 and illustrated in Figure 6. Overall, the cumulative changes result in nearly doubled capital investment and first-year revenue requirements. The investment increases resulting from the new economic premises are related to higher indirect capital investment costs, particularly in interest during construction, contractor expense, and working capital. The increase in first-year revenue requirements stems largely from capital charges based on the capital investment. New power plant coal and air rates, the operating profile, and the 1979 NSPS all produce similar increases in capital investment. In these cases the main factors are increased flue gas volume, increased lifetime waste disposal requirements, and the more stringent scrubbing conditions. The effect on annual revenue requirements is similar except, of course, that the reduction in yearly operating hours results in a reduction in costs. Addition of reliability factors (a spare scrubber train, emergency bypass, and a spare ball mill) also cause appreciable increases in both capital investment and first-year

TABLE 5. COMPARISON OF OLD AND NEW PREMISE CONDITIONS USING THE LIMESTONE SCRUBBING PROCESS

	Old premises	New premises
Design Premises		
Coal, Btu/lb	10,500	11,700
Excess air, %	33	39
Heat rate, Btu/kWh	9,000	9,500
Operating profile		
First year, hr/yr	7,000	5,500
Lifetime, hr (30 years)	127,000	165,000
FGD		
SO_x removal, %	1971 NSPS	1979 NSPS
Emergency bypass, %	0	50
Spare units	0	1
Booster fan	FD	ID
Limestone process		
Absorber	Mobile bed	Spray tower
L/G, gal/kaft^3	50	90
Gas velocity, ft/sec	12.5	10.0
ΔP, in. H_2O	8	1.4
Forced oxidation	No	Yes
Waste disposal	Pond	Landfill
Economic Premises		
Cost index year		
Capital investment	1979	1982
Annual revenue requirements	1980	1984
Indirect capital costs		Revised
Land, $/acre	3,500	5,000
Interest during construction, %	12	15.6
Limestone process contingency, %	20	10
Pond contingency, %	20	10
Pond allowance for startup, %	8	0
Capital charges	Average annual	Levelized
Depreciation	Straight line	Sinking fund

TABLE 6. COST COMPARISON IN TRANSITION FROM OLD TO NEW PREMISES AND TECHNOLOGY FOR THE LIMESTONE SCRUBBING PROCESS

Condition	Capital investment				First-year revenue requirements			
	k$	$/kW	% change	% total change	k$	Mills/kWh	% change	% total change
Old premises and technology	48,700	98	-	-	14,100	4.0	-	-
Above with new economic premises and pond	55,100	110	13	13	16,200	4.6	15	15
Above with new power plant design premises	57,100	114	4	17	17,000	4.9	5	21
Above with new operating profile	59,800	120	5	23	16,500	6.0	-3	17
Above with 1979 NSPS	63,600	127	6	29	17,200	6.3	4	22
Above with reliability factors (spares and bypass)	77,100	154	21	58	20,100	7.3	17	43
Above with spray tower	83,300	167	8	71	21,500	7.8	7	52
Above with landfill	76,000	152	-9	56	21,700	7.9	1	54
Above with 1982, 1984 costs	96,800	194	28	99	27,300	9.9	26	94

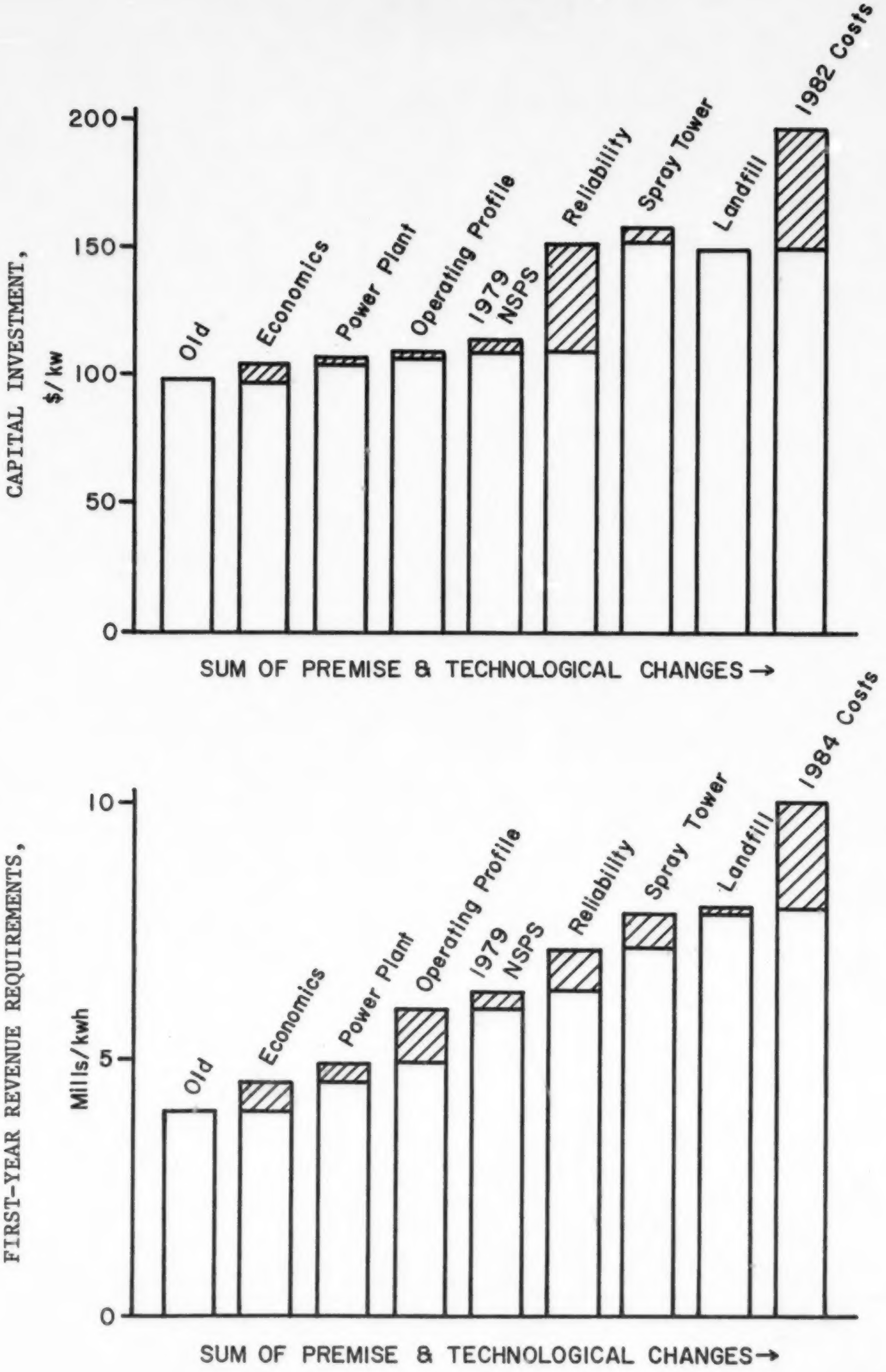


Figure 6. Stepwise conversion of limestone scrubbing costs from old to new premises and technology.

revenue requirements. The use of a spray tower instead of a mobile-bed absorber increases costs primarily because of the lower flue gas velocity and higher slurry recirculation rate, which requires larger ducting and pumping requirements.

Substitution of landfill for ponding substantially reduces capital investment by eliminating pond construction costs. The resulting reduction in capital charges essentially counteracts the increased waste disposal costs in first-year revenue requirements.

The largest cost increase is a result of advancing the cost index year from 1978 to 1982 for capital investment and from 1980 to 1984 for first-year revenue requirements.

Overall, economics in the form of inflation and higher interest have the largest effect in comparison of the limestone process using the old and new premises and technology. Technical changes related to improvements in reliability, such as bypass and redundancy provisions, also have a large effect. The higher SO_x removal efficiency has less effect than the economic and technical changes.

ADVANCED LIMESTONE SCRUBBING TECHNOLOGY

As stated earlier, TVA is now conducting an EPA-sponsored economic evaluation of advanced limestone scrubbing technology. The study encompasses recent developments in limestone scrubbing such as chemical additives, increasing use of spray towers, forced oxidation, and landfill techniques. The complete results of this project will be published in 1981.

Of particular interest at this time is the advanced limestone system using a spray tower, forced oxidation, adipic acid addition and landfill of the gypsum waste. The interest comes from favorable results at the Shawnee Test Facility. Earlier bench- and pilot-scale studies were made by TVA and EPA on adipic acid addition and EPA is sponsoring an adipic acid demonstration unit at the Southwest Plant of Springfield (Missouri) City Utilities. The advantage of adipic acid (or other similar additives) lies in its buffering action, which controls the slurry pH at more favorable reaction conditions. This increases the reactivity of the slurry, improving SO_2 removal efficiency and increasing limestone utilization.

As a special feature, an economic comparison of the advanced process with the new conventional and old conventional limestone processes is in order. The design conditions for the three processes are shown in Table 7.

Tables 8 and 9 show the capital investments and annual revenue requirements for the three processes based on the base-case conditions and the new premises that were discussed previously. The costs thus

TABLE 7. PROCESS DESIGN CONDITIONS AND PREMISES - LIMESTONE PROCESSES

	Advanced process	New conventional	Old conventional
Type of absorber	Spray tower	Spray tower	Mobile bed
Forced oxidation	Yes	Yes	No
Adipic acid use	Yes (1000 ppm)	No	No
Waste disposal	Thickener-filter-landfill	Thickener-filter-landfill	Pond
Scrubber gas velocity, ft/sec	10	10	12.5
L/G, gal/kaft^3	80	90	58
Limestone stoichiometry	1.2	1.3	1.3
Air stoichiometry	2.5	2.5	0
Percent sulfite oxidation	95	95	30
ID fan/FD fan	ID	ID	ID
Spare scrubber	Yes	Yes	Yes
Filter cake solids, %	80	80	-
Pond settled solids, %	-	-	40
Spare ball mill	Yes	Yes	Yes
Reheat	In-line steam	In-line steam	In-line steam
Bypass available	50% emergency	50% emergency	50% emergency

TABLE 8. CONVENTIONAL AND ADVANCED LIMESTONE SCRUBBING PROCESSES

CAPITAL INVESTMENT

	Capital investment, k$		
	Old conventional[a]	New conventional[b]	Advanced[c]
Direct Investment			
Material handling	3,498	3,497	3,503
Feed preparation	3,485	3,484	3,490
Gas handling	9,600	11,129	10,821
SO_2 absorption	19,830	22,988	22,351
Reheat	2,851	3,304	3,213
Solids disposal	2,063	2,868	2,850
Total	41,327	47,270	46,228
Services, utilities, and miscellaneous	2,480	2,836	2,774
Total	43,807	50,106	49,002
Landfill or pond construction	13,960	2,076	1,983
Landfill equipment	-	500	495
Total	57,767	52,682	51,480
Indirect Investment			
Engineering design and supervision	3,346	3,663	3,579
Architect and engineering contractor	1,016	1,028	1,005
Construction expense	8,126	8,378	8,187
Contractor fees	2,888	2,608	2,549
Contingency	7,315	7,158	6,990
Total fixed investment	80,458	75,517	73,790
Other Capital Investment			
Allowance for startup and modifications	5,012	5,732	5,606
Interest during construction	12,551	11,781	11,511
Land	1,905	641	611
Working capital	3,104	3,161	3,090
Total capital investment	103,030	96,832	94,608
$/kW	206	194	189

Basis

Upper Midwest plant location represents project beginning mid-1980, ending mid-1983. Average cost basis, mid-1982. Spare pumps, one spare scrubbing train, and one spare ball mill are included. Disposal pond and landfill located 1 mile from plant. Investment includes FGD feed plenum but excludes stack plenum and stack.

a. Old conventional process is a mobile bed absorber with onsite ponding of sulfite sludge.

b. New conventional process is a spray tower, forced oxidation and gypsum landfill.

c. Advanced system is same as b. but with adipic acid addition for enhanced SO_2 removal.

TABLE 9. CONVENTIONAL AND ADVANCED LIMESTONE SCRUBBING PROCESSES

ANNUAL REVENUE REQUIREMENTS

	Annual cost, k$		
	Old conventional	New conventional	Advanced
Direct Costs - First-Year			
Raw materials			
Limestone	1,128	1,128	1,041
Adipic acid	-	-	216
Total raw materials cost	1,128	1,128	1,257
Conversion costs			
Operating labor and supervision			
FGD	460	658	658
Solids disposal	-	529	517
Utilities			
Process water	35	26	26
Electricity	1,732	2,018	1,874
Steam	1,273	1,365	1,367
Fuel	-	199	189
Maintenance			
Labor and material	3,923	4,025	3,937
Analyses	104	104	104
Total conversion costs	7,527	8,924	8,672
Total direct costs	8,655	10,052	9,929
Indirect Costs - First-Year			
Overheads			
Plant and administrative (60% of conversion costs less utilities)	2,692	3,057	2,998
Total first-year operating and maintenance costs	11,347	13,109	12,927
Levelized capital charges (14.7% of total capital investment)	15,145	14,234	13,907
Total first-year annual revenue requirements	26,492	27,343	26,834
Levelized first-year operating and maintenance costs (1.886 x first-year O and M)	21,401	24,724	24,381
Levelized capital charges (14.7% of total capital investment)	15,145	14,234	13,907
Levelized annual revenue requirements	36,545	38,958	38,288
	Mills/kWh		
First-year annual revenue requirements	9.63	9.94	9.76
Levelized annual revenue requirements	13.29	14.17	13.92

Basis
Upper Midwest plant location, 1984 revenue requirements.
New plant with 30-year life.
Power unit on-stream time, 5,500 hr/yr.
Coal burned, 1,116,500 tons/yr.
Boiler heat rate, 9,500 Btu/kWh.
Total capital investment:
Old conventional - $103,030,000
New conventional - $ 96,832,000
Advanced - $ 94,608,000

incorporate a spare scrubber, emergency bypass, and a 1981-1983, 1984 time period, among other differences from the FGD studies discussed previously. All of the costs except those for landfill were developed by the TVA Shawnee Computer Economics Program (15).

Both the new conventional process and the advanced process have slightly higher direct capital investment costs than the old conventional process in most areas. The old conventional process has disposal site (pond) construction costs over ten times higher than the disposal site (landfill) construction costs than the others, however. The result is a slightly lower capital investment for the new conventional and advanced processes. The use of adipic acid in the advanced process produces a minor increase in material handling costs and much larger decreases in absorber and disposal costs. The increased reactivity of the limestone slurry allows both less stringent scrubbing conditions and improved limestone utilization, resulting in lower limestone consumption and less unreacted limestone in the waste.

In annual revenue requirements, the old conventional process has lower conversion costs, primarily because of lower labor and supervision and electricity costs, resulting in lower overall expense. The increase in labor and supervision cost for the new conventional and advanced processes is largely for disposal operations because trucking and earth-moving operations are required. In comparison of the new conventional process and the advanced process, adipic acid addition causes a slight overall reduction in costs, primarily because of lower limestone and electricity consumption.

REFERENCES

1. S. V. Tomlinson, F. M. Kennedy, F. A. Sudhoff, and R. L. Torstrick. Definitive SO_x Control Process Evaluations - Limestone, Double-Alkali, and Citrate FGD Processes. TVA ECDP B-4, Tennessee Valley Authority, Office of Power, Emission Control Development Projects, Muscle Shoals, Alabama. EPA-600/7-79-177, U.S. Environmental Protection Agency, Office of Research and Development, Washington, D.C., 1979.

2. K. D. Anderson, J. W. Barrier, W. E. O'Brien, and S. V. Tomlinson. Definitive SO_x Control Process Evaluations: Limestone, Lime, and Magnesia FGD Processes. TVA ECDP B-7, Tennessee Valley Authority, Office of Power, Emission Control Development Projects, Muscle Shoals, Alabama. EPA-600/7-80-001, U.S. Environmental Protection Agency, Office of Research and Development, Washington, D.C., 1980.

3. J. R. Byrd, K. D. Anderson, S. V. Tomlinson, and R. L. Torstrick. Definitive SO_x Control Process Evaluations: Aqueous Carbonate, Wellman-Lord, Allied Chemical, and Resox® FGD Technologies. Tennessee Valley Authority, Office of Power, Division of Energy Demonstrations and Technology, Muscle Shoals, Alabama. U.S. Environmental Protection Agency, Office of Research and Development, Washington, D.C. (In press)

4. G. G. McGlamery, R. L. Torstrick, J. W. Broadfoot, J. P. Simpson, L. J. Henson, S. V. Tomlinson, and J. F. Young. Detailed Cost Estimates for Advanced Effluent Desulfurization Processes. TVA Bulletin Y-90, Tennessee Valley Authority, Office of Agricultural and Chemical Development, Muscle Shoals, Alabama. EPA-600/2-75-006, U.S. Environmental Protection Agency, Office of Research and Development, Washington, D.C., 1975.

5. J. W. Barrier, H. L. Faucett, and L. J. Henson. Economics of Disposal of Lime-Limestone Scrubbing Wastes: Untreated and Chemically Treated Wastes. TVA Bulletin Y-123, Tennessee Valley Authority, National Fertilizer Development Center, Muscle Shoals, Alabama. EPA-600/7-78-023a, U.S. Environmental Protection Agency, Office of Research and Development, Washington, D.C., 1978.

6. J. W. Barrier, H. L. Faucett, and L. J. Henson. Economics of Disposal of Lime/Limestone Scrubbing Wastes: Sludge/Flyash Blending and Gypsum Systems. TVA Bulletin Y-140, Tennessee Valley Authority, National Fertilizer Development Center, Muscle Shoals, Alabama. EPA-600/7-79-069, U.S. Environmental Protection Agency, Office of Research and Development, Washington, D.C., 1979.

7. J. D. Veitch, A. E. Steele, and T. W. Tarkington. Economics of Disposal of Lime/Limestone Scrubbing Wastes: Surface Mine Disposal and Dravo Landfill Processes. TVA EDT-105, Tennessee Valley Authority, Office of Power, Division of Energy Demonstrations and Technology, Muscle Shoals, Alabama. EPA-600/7-80-022, U.S. Environmental Protection Agency, Office of Research and Development, Washington, D.C., 1980.

8. J. I. Bucy, J. L. Nevins, P. A. Corrigan, and A. G. Melicks. The Potential Abatement Production and Marketing of Byproduct Elemental Sulfur and Sulfuric Acid in the United States. TVA S-469, Tennessee Valley Authority, Office of Agricultural and Chemical Development, Muscle Shoals, Alabama, 1976.

9. J. I. Bucy, R. L. Torstrick, W. L. Anders, J. L. Nevins, and P. A. Corrigan. Potential Abatement Production and Marketing of Byproduct Sulfuric Acid in the U.S. TVA Bulletin Y-122, Tennessee Valley Authority, Office of Agricultural and Chemical Development, Muscle Shoals, Alabama. EPA-600/7-78-070, U.S. Environmental Protection Agency, Washington, D.C., 1978.

10. W. E. O'Brien and W. L. Anders. Potential Production and Marketing of FGD Byproduct Sulfur and Sulfuric Acid in the U.S. (1983 Projection). ECDP B-1, Tennessee Valley Authority, Office of Power, Emission Control Development Projects, Muscle Shoals, Alabama EPA-600/7-79-106, U.S. Environmental Protection Agency, Washington, D.C., 1979.

11. W. L. Anders. Computerized FGD Byproduct Production and Marketing System: Users Manual. TVA ECDP B-2, Tennessee Valley Authority, Office of Power, Emission Control Development Projects, Muscle Shoals, Alabama. EPA-600/7-79-114, U.S. Environmental Protection Agency, Washington, D.C., 1979.

12. W. E. O'Brien, W. L. Anders, and J. D. Veitch. Projection of 1985 Market Potential for FGD Byproduct Sulfur and Sulfuric Acid in the U.S. TVA EDT-115, Tennessee Valley Authority, Office of Power, Division of Energy Demonstrations and Technology, Muscle Shoals, Alabama. EPA-600/7-80-131, U.S. Environmental Protection Agency, Washington, D.C., 1980.

13. T. A. Burnett and W. E. O'Brien. Preliminary Economic Analysis of a Lime Spray Dryer FGD System. TVA EDT-112, Tennessee Valley Authority, Office of Power, Division of Energy Demonstrations and Technology, Muscle Shoals, Alabama. EPA-600/7-80-050, U.S. Environmental Protection Agency, Washington, D.C., 1980.

14. T. A. Burnett and K. D. Anderson. Technical Review and Economic Evaluation of Spray Dryer FGD Systems. Tennessee Valley Authority, Office of Power, Division of Energy Demonstrations and Technology, Muscle Shoals, Alabama. U.S. Environmental Protection Agency, Washington, D.C. (In press)

15. C. D. Stephenson and R. L. Torstrick. Shawnee Lime/Limestone Scrubbing Computerized Design/Cost-Estimate Model Users Manual. ECDP B-3, Tennessee Valley Authority, Office of Power, Emission Control Development Projects, Muscle Shoals, Alabama. EPA-600/7-79-210, U.S. Environmental Protection Agency, Washington, D.C., 1979.

SO_2 AND NOx ABATEMENT FOR COAL-FIRED BOILERS IN JAPAN

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The total capacity of coal-fired utility boilers in Japan, which was only 4,300 MW (3.7% of total utility power) in 1979, is expected to increase to 10,000 MW (5.6%) in 1885 and to 22,000 MW (10.0%) in 1990. Most of the boilers will apply FGD by the wet limestone-gypsum process because of its reliability and relatively low cost. To save energy and water, FGD systems with a low pressure drop and small water consumption are preferred. Tests on FGD by a dry carbon process are under way.

NOx concentrations in flue gases from existing coal-fired boilers have been lowered to 200 - 350 ppm by combustion modification including staged combustion and the use of low-NOx burners. For further abatement, selective catalytic reduction (SCR) has started to be applied to several coal-fired boilers. The first full-scale combination system of SCR and FGD was put into operation in April 1980. The plant cost for SCR is about one-third that for FGD. A new combustion technology has also been developed in attempts to lower NOx below 100 ppm.

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1. COAL USAGE AND POLLUTION CONTROL IN JAPAN

Most utility power companies in Japan switched fuel from coal to oil between 1960 and 1974 except Electric Power Development Co. (EPDC) which was established by the Japanese government jointly with major power companies to use domestic coal. Due to the recent rise in oil and gas prices, power companies have started to construct new coal-fired boilers (Table 1), most of which will use imported coal because the supply of domestic coal is limited to 20 million tons yearly. Although Japan has imported over 60 million tons of coal yearly, all of the imported coal has been for coke production for the steel industry. The import of fuel coal has been started and is expected to reach 45 million tons in 1990.

Major problems with coal usage are (1) emissions of SO_2, NOx and particulates on combustion, (2) handling and storage problems, and (3) ash disposal. Those problems are serious in Japan where a large population is concentrated in a small land space. The new boilers are to be located in regions relatively far from large cities and industrial districts, where the environmental regulations by the Central Government are not quite stringent. However, in order to construct a large plant, it is necessary to make an agreement with local governments, by which extensive countermeasures for pollution control are necessitated.

All of the new coal-fired boilers will need FGD. NOx concentrations in flue gas from major coal-fired boilers has been reduced to 200 - 350 ppm while the emission standard by the Central Government is 400 ppm for new boilers and 480 ppm for existing ones. Further reduction will be needed for new boilers. Some power companies have started to apply selective catalytic reduction (SCR) which usually removes about 80% of NOx (Table 1).

A new combustion technology to lower NOx concentration below 100 ppm with coal and below 50 ppm with oil has been developed. (Section 6.2).

Particulates can be removed sufficiently by a combination of electrostatic precipitator and wet FGD. A bag house has been tested but has not been considered promising for a large boiler.

In attempts to solve the coal handling problem, coal-oil mixture (COM) has been studied extensively and may be used for some of the new boilers. The major drawback with COM is that more than half of the energy is derived from oil. To save oil, coarse-grain COM has been tested, which uses up to 6 mm grains of coal which is transported with oil as a slurry and separated from oil for burning.

The largest problem with coal usage may be ash disposal, because landspace for discarding is limited. New uses of the ash, as feedstock

Table 1 Coal-fired utility boilers in Japan

(Larger than 175 MW)

Power company	Power station	Boiler No.	Capacity MW	Year of Completion Boiler	Year of Completion FGD	Year of Completion SCR[a]
EPDC	Isogo	1	265	1967	1976	
		2	265	1969	1976	
"	Takasago	1	250	1968	1975	
		2	250	1969	1976	
"	Takehara	1	250	1967	1977	1981
		3	700	1982	1982	1982
"	Matsushima	1	500	1981	1981	
		2	500	1981	1981	
"	Matsuura	1	1,000	1984[b]		
		2	1,000	1986[b]		
"	Mito	1	1,000	1988[b]		
Chugoku	Shimonoseki	1	175	1967	1979	1980
"	Misumi	1	700	1985[b]		
Hokkaido	Tomato-Atsuma	1	350	1980	1980	1980[c]
"	Sunagawa	4	125	1982	1982	
Kyushu	Matsuura	1	700	1984[b]		
		2	700	1988[b]		
"	Reihoku	1	700	1987[b]		
		2	700	1989[b]		
Joban Kyodo	Nakoso	8	600	1983[d]		1983
		9	600	1983[d]		1983
Tohoku	Noshiro	1	600	1985[b]		
		2	600	1985[b]		
Tokyo	Mito	1	1,000	1988[b]		
		2	1,000	1988[b]		

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[a] Selective catalytic reduction of NOx

[b] Planned.

[c] Treating one-fourth of the gas.

[d] Mostly oil will be used with less coal for a while without FGD.

for cement production replacing clay, as filler for asphalt, as raw material for aggregate, etc., have been developed.

Studies have been carried out also on fluidized bed combustion (FBC), gasification, and liquefaction of coal, but not as extensively as in the USA. The major problem with FBC in Japan is the difficulty in disposing of the ash containing lime and calcium sulfate. Tests have been conducted in search for an SO_2 absorbent that can be separated from ash, regenerated and recycled, but so far do not seem promising. Gasification and liquefaction may not be suitable to Japan which has to depend on imported coal, since a considerable portion of energy of coal is consumed by gasification or liquefaction. Although liquefaction may be important in future, the plant may have to be constructed abroad and the product imported.

2. STATUS OF FGD FOR COAL-FIRED UTILITY BOILERS

Before 1979, FGD plants for coal-fired utility boilers were limited to the 5 plants of EPDC. Among the EPDC plants, two at Takasago Station had an appreciable scaling problem until 1977 mainly at the mist eliminator which had been washed with a circulating liquor saturated with gypsum.1) By using fresh water together with the liquor for the wash, the scaling problem was solved.1,2) Since 1978, all of EPDC's FGD plants have been operated with virtually 100% operability and reliability (Table 2).

Table 2 Operation hours of EPDC's boilers and FGD plants
(April 1978 through March 1979)

Boiler		Operation hours Boiler (A)*	FGD (B)	B/A (%)
Isogo	No. 1	7,705	7,705	100.0
	No. 2	8,206	8,206	100.0
Takasago	No. 1	7,829	7,823	99.92
	No. 2	8,167	8,147	99.75
Takehara	No. 1	7,583	7,580	99.95

* When an FGD plant is shut down due to its own trouble, the boiler is operated by using low-sulfur oil. Therefore, B/A (%) shows operability as well as reliability.

Operation parameters of the plants are shown in Table 3. Although the plants are highly reliable and removes over 90% of SO_2 and over 70% of fly ash, they have the following drawbacks: (1) A large gas pressure drop due to the use of a venturi or perforated plate scrubber to attain a high dust removal efficiency, which results in a large power consumption. (2) Requirement of a large amount of water for gas cooling and also for purging wastewater from the system in order to maintain chlorine in the scrubber liquor below a certain level for corrosion prevention. (Usually more than half of the water charged to the FGD system is volatilized in the prescrubber).

In order to lower the pressure drop, new FGD plants, including Chugoku Electric's Shimonoseki plant constructed by MHI and two EPDC plants at Matsushima under construction by Babcock Hitachi and IHI, use a spray tower for gas cooling and particulate removal. A gas-gas heater (heat exchanger) is used for the new plants as well as the Tomato-Atsuma plant of Hokkaido Electric in order to cool the FGD inlet gas to save water and to heat the FGD outlet for energy conservation.

Dry processes for FGD have received attention as a possible way for further improvement and also because of the convenience for use in conjunction with selective catalytic reduction of NOx. An activated carbon process has been tested at EPDC's Takehara station. (Section 6.1). The Electric Power Industry Federation also is to make pilot plant tests on activated carbon processes for coal-fired boilers at 3 power stations.

3. NOx ABATEMENT AND COMBINATION OF SCR AND FGD

3.1 NOx Regulation and Selective Catalytic Reduction (SCR)

NOx concentration in flue gases from coal-fired boilers has been restricted by the emission standards by the central government to a level below 480 ppm for existing boilers and below 400 ppm for new boilers. The concentration can be achieved by combustion modification without appreciable difficulty. Most local governments, however, enforce much more stringent regulations. For example, Yokohama City, in an effort to lower the ambient NO_2 concentration from the current 0.06 - 0.07 ppm in daily average to 0.04 ppm, has asked EPDC's Isogo Station to lower to 169 ppm the NOx concentration in flue gases from the existing two 265 MW coal-fired boilers. EPDC has lowered the NOx concentration to 200 ppm by combustion modification including staged combustion and low-NOx burner and has been making further efforts to meet the requirement. Isogo Station has a limited landspace in which they managed to retrofit FGD plants and has no more space to install a flue gas treatment (FGT) plant for NOx removal. Therefore,

Table 3 Operation parameters of FGD plants for coal-fired utility boilers

Power company	EPDC	EPDC	EPDC	EPDC		Chugoku	Hokkaido
Station	Isogo	Takasago	Takehara	Matsushima		Shimonoseki	Tomato
Boiler No.	1	1	1	1	2	1	1
Boiler capacity (MW)	265	250	250	500	500	175	350
FGD constructor	IHI[a]	Mitsui[b]	BH[c]	IHI[a]	BH[c]	MHI[d]	BH[c]
FGD start-up	May '76	Feb. '75	Feb. '77	Jan. '81	Jan. '81	July '79	Oct. '80
Gas treated (1,000 Nm^3/hr)	821	792	793	1,826	1,826	586	1,268
Inlet SO_2 (ppm)	450	1,500	1,730	1,000	1,000	1,310	232
Inlet dust (mg/Nm^3)	1,500	100	400	300	300	830	45
Prescrubber (cooler)							
Type	Venturi	Venturi	Venturi	Spray	Spray	Spray	Venturi
L/G (liters/Nm^3)	7	6	2.5	2.8		3	
Scrubber (SO_2 absorber)							
Type	Venturi	Venturi	PP[e]	Spray	Spray	Packed	PP[e]
L/G	7	6	7	13.4	15	14	
Outlet SO_2 (ppm)	25	100	100	50	50	55	23
Outlet dust (mg/Nm^3)	50	30	50	30	30	50	
SO_2 removal efficiency (%)	94.4	93.3	94.2	95.0	95.0	95.8	90.0
Dust removal efficiency (%)	96.6	70.0	87.5	90.0	90.0	94.0	
Pressure drop (mm H_2O)	360[f]	325[f]	615[f]		133[g]	120[f]	
Wastewater (t/hr)	10	7.5	12			15	
Energy requirement (%)[h]	2.9	3.2	3.3			2.1	
Reliability (%)[i]	100.0	99.9	100.0			100.0	

[a] Ishikawajima-Harima Heavy Industries

[b] Mitsui Miike Machinery Co.

[c] Babcock Hitachi K.K.

[d] Mitsubishi Heavy Industries

[e] Perforated plate

[f] By two scrubbers and mist eliminators

[g] By two scrubbers

[h] Percent of power generated

[i] EGD operation hours percent of desired FGD operation hours

they need to reduce NOx further by improved combustion. Even more stringent regulations may be applied for new larger boilers, necessitating FGT.

Among many ways of FGT developed in Japan, selective catalytic reduction (SCR) that uses ammonia and catalyst at 300 - 400°C is by far the most advanced method, which has been used in constructing about 100 commercial plants mainly for flue gas from oil-fired boilers. The advantages of SCR over other FGT processes are simplicity and reliability which enable unattended operation, lack of the by-product disposal problem, and relatively low cost. SCR is conveniently applied to flue gas leaving a boiler economizer at 300 - 400°C. The major reaction is shown below:

$$4NO + 4NH_3 + O_2 = 4N_2 + 6H_2O$$

At the early stage of development, SCR encountered the following technical problems, most of which have been solved by recent improvements: (1) Catalyst poisoning by SOx in flue gas. (2) Catalyst pluggage by dust. (3) Catalytic oxidation of a portion of SO_2 to SO_3. (4) Leak ammonia from SCR reactor, which reacts with SO_3 and H_2O to form ammonium bisulfate deposit in an air preheater.

Many of the catalysts developed recently are based on TiO_2 with small amounts of V_2O_5 and other components, are resistant to SOx, and oxidize about 1% or less SO_2. In order to prevent dust plugging of the catalyst, parallel flow type reactors with honeycomb, plate, and tube catalysts have been used for dusty gases such as coal-fired boiler flue gas.

More than 90% of NOx can be removed by using over 1 mol NH_3 to 1 mol NOx as shown in Figure 1. However, 80% removal has been generally applied to utility boilers as the optimum control level, because compared with 90% removal, it requires about 40% less catalyst resulting in the reduction of cost as well as pressure drop and also because it can reduce leak ammonia to a low level (5 ppm or below) to minimize the deposit of ammonium bisulfate in the air preheater. Over 90% removal with a low leak NH_3 is difficult for a large boiler because the gas velocity as well as NOx concentration is not uniform in different parts of the duct.

Low-temperature catalysts active at 150 - 250°C have also been developed but have not been used commercially yet because ammonium bisulfate forms on the catalyst and lowers its activity. Ammonium bisulfate can be removed by heating the catalyst to over 350°C. The low-temperature catalyst may not be suitable for boilers for which economizer outlet gas around 350°C can be treated but may be useful for other sources for which only cold gas around 200°C is available.

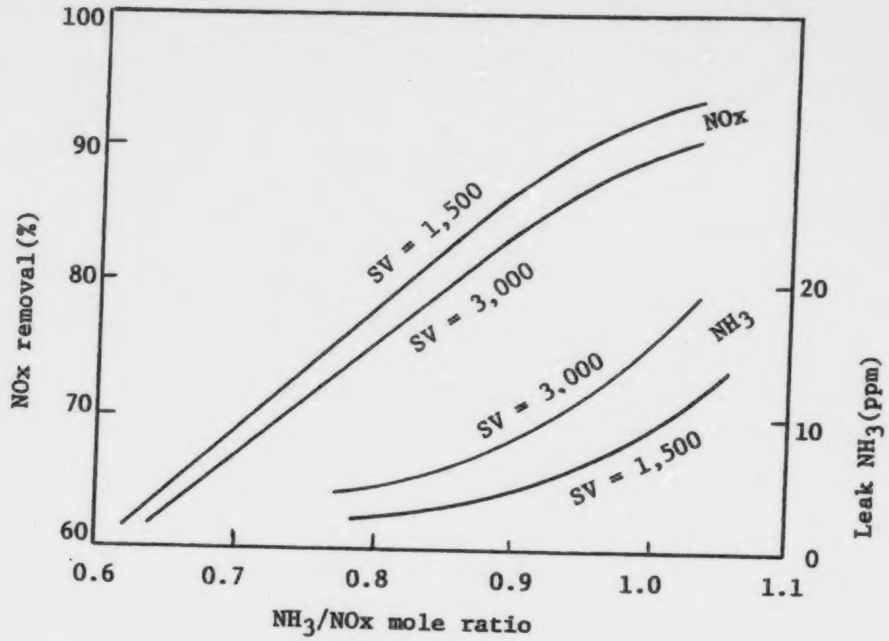
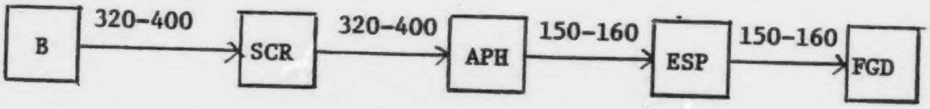


Figure 1 Performance of honeycomb catalyst for coal-fired boiler flue gas (Inlet NOx 300 ppm, at 370°C. SV means space velocity: flue gas volume per hour divided by catalyst volume. For high-dust system)

High-dust system


B
320-400
SCR
320-400
APH
150-160
ESP
150-160
FGD

Low-dust system

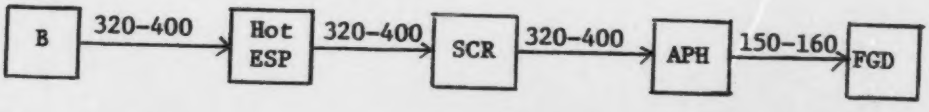

B
320-400
Hot
ESP
320-400
SCR
320-400
APH
150-160
FGD

Figure 2 Systems for coal-fired boiler flue gas treatent (Figures show gas temperature, °C. B: Boiler APH: Air preheater)

3.2 Combination of SCR and FGD

At an early stage of development, the SCR reactor was placed downstream of FGD in order to reduce SOx poisoning and dust plugging. This system, however, requires a large amount of energy for heating the gas after FGD and has not been used since SOx-resistant parallel flow type catalysts have been developed. Figure 2 shows two combination SCR/FGD systems currently used for coal-fired boilers. In both systems, the economizer outlet gas at 330 - 400°C is treated by SCR, cooled to 150°C by an air preheater, and then subjected to FGD. The high dust system treats the gas with full dust load (15 - 25 grams/Nm^3) by SCR, and therefore the catalyst should be hard in order to avoid erosion by dust and thus is less porous and may not be highly active. On the other hand, the low dust system uses a hot electrostatic precipitator (ESP) upstream of SCR, which is suitable for dust removal in flue gas from low-sulfur coal. The hot ESP usually reduces the dust to 100 - 200 mg/Nm^3 and protects the catalyst from erosion. However, the dust leaving the hot ESP is finer and richer in alkaline components and tends to deposit on the catalyst surface. The problem of ammonium bisulfate deposit in the air preheater is also appreciable with the low dust system while it is insignificant with the high dust system (Section 5.3). Therefore, leak ammonia should be kept at a lower level with the low dust system than with the high dust system.

As shown in Table 4, the Shimonoseki plant, Chugoku Electric uses the high dust system while the Tomato-Atsuma plant of Hokkaido Electric and the plants at Takehara, EPDC use the low-dust system. Two plants at Nakoso, Joban Joint Electric will use the high dust system.

Table 4 SCR plants for coal-fired utility boilers

Company	Station	Capacity (MW)	Vendor	SCR type Dust	SCR type Catalyst	Completion
Chugoku	Shimonoseki	175	MHI	High	Honeycomb	1980
Hokkaido	Tomato-Atsuma	350 x 1/4	BH	Low	Plate	1980
EPDC	Takehara	250 x 1/2	BH	Low	Plate	1981
		250 x 1/2	KHI[a]	Low	Tube	1981
EPDC	Takehara	700	nd[b]	Low	nd[b]	1982
Joban	Nakoso	600	MHI	High	Honeycomb	1983
		600	IHI	High	Honeycomb	1983

a Kawasaki Heavy Industries b Not decided

The flue gas leaving the SCR reactor contains a small amount of ammonia, which is caught by a prescrubber of the FGD system. Although ammonia has no adverse effect either on the operation of wet lime/limestone process FGD or on the quality of by-product gypsum, it is contained in a small amount in wastewater from the FGD system. If needed, the ammonia in the wastewater can be removed by a conventional biochemical treatment (activated sludge process) or by ammonia stripping. The latter has been used at the Owase plant, Chubu Electric while the former is to be used at the Takehara plant, EPDC.

3.3 SCR Cost

Examples of SCR plant cost for utility boilers are shown in Table 5. The cost for the new gas-fired boiler at Chita was 1,860 yen/kW, while that for the new oil-fired boiler at Kudamatsu was 2,860 yen/kW. Those for existing oil-fired boilers at Kudamatsu and Chita were considerably higher than that for the new oil-fired boiler, because of complicated duct work for retrofitting (Kudamatsu and Chita) and the requirement of additional fans (Kudamatsu). The SCR plant for coal at Shimonoseki is more costly than for oil.

The difference in cost with the fuel type is due mainly to the amount of catalyst needed. Generally speaking, an active pellet catalyst can be used for clean gas, while for flue gas from oil containing 20 - 100 mg/Nm^3 of dust, a honeycomb catalyst with a channel size of 6 - 7 mm and wall thickness of 1 - 1.5 mm consisting of SOx resistant material has been used in a volume 3 - 4 times that of the pellet catalyst. For coal, the catalyst volume may be nearly double that for oil because of a larger channel size of honeycomb for dust plugging prevention and a harder structure for erosion prevention resulting in lower activity.

Estimated SCR costs for new 700 MW utility boilers using coal and low-sulfur oil are shown in Table 6. Honeycomb catalyst is used for both oil and coal. The assumed channel size and wall thickness in millimeters are 6.6 and 1.4 for oil, 7.4 and 1.6 for coal with the low-dust system, and 8.2 and 1.8 for coal with the high-dust system. Leak ammonia is maintained below 10 ppm for oil (low sulfur) and coal with the high dust system while it is kept below 5 ppm for coal with the low-dust system which is liable to air preheater plugging. Based on those assumptions, an equal space velocity was assumed for high and low dust systems of coal. The space velocity is about one-half that for oil.

The investment cost including civil engineering and test operation for 80% NOx removal is nearly 4,000 yen/kW for oil and nearly 7,000 yen/kW for coal, while the cost for 90% removal is higher by about 30% for oil and 40% for coal. The annualized SCR costs in yen/kWhr for 80% removal, assuming 7 years' depreciation, 70% boiler utilization,

Table 5 Cost of SCR plants for utility boilers (in battery limits)

Power company	Power station	Boiler (MW)	Fuel	New or retrofit	NOx removal (%)	Const-ructor	Catalyst type	Space velocity[a] (hr^{-1})	Plant cost 10^9yen	Plant cost yen/kW	Year com-pleted
Chubu	Chita	700	Gas	New	Over 80	BH	Pellet	20,000	1.3	1,860	1977
Chubu	Chita	700	Oil[b]	Retrofit	Over 80	MHI	Honeycomb	6,000	2.2	3,570	1980
Chugoku	Kudamatsu	375	Oil[b]	Retrofit	Over 80	IHI	Honeycomb	5,500	2.2	5,870	1979
Chugoku	Kudamatsu	700	Oil[b]	New	Over 80	IHI	Honeycomb	5,500	2.0	2,860	1979
Chugoku	Shimonoseki	175	Coal	Retrofit	Over 50[c]	MHI	Honeycomb	3,000	1.7[d]	9,710[d]	1980

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[a] Flue gas volume per hour divided by catalyst volume

[b] Low-sulfur oil

[c] Catalyst for 50% removal has been used to meet the current regulation, while the SCR system has been designed for 80% removal.

[d] Including boiler modification for economizer bypass.

Table 6 Estimated SCR cost for new 700 MW utility boilers

Annual power generation 4,292,400 MWhr. 70% utilization.
Leak NH_3: 5 - 10 ppm for oil and coal with high-dust system. Less than 5 ppm for coal with low-dust system

Fuel	Oil (low S)		Coal (high and low dust)	
Flue gas, Nm^3/hr. (NOx ppm)	2,000,000 (120)		2,300,000 (300)	
NOx removal efficiency (%)	80	90	80	90
Space velocity (hr^{-1})	5,100	3,400	2,700	1,700
Investment cost				
Catalyst (billions of yen)[a]	1.22	1.82	2.81	4.46
Other (")[b]	1.50	1.75	2.00	2.30
Total (")[b]	2.72	3.63	4.81	6.76
Total (1,000 yen/kW)	3.89	5.10	6.87	9.66
Annual cost (billions of yen)				
Capital cost[c]	0.50	0.62	0.78	1.02
Catalyst[d]	0.61	0.91	2.81	4.46
Other[e]	0.27	0.31	0.48	0.55
Total	1.38	1.84	4.07	6.03
Annualized cost (yen/kWhr)	0.32	0.43	0.95	1.40
(1,000 yen/Nm^3 of NOx removed)	1.15	1.39	1.20	1.58

---

[a] 3.1 million yen/m^3 for oil, 3.3 million yen/m^3 for coal.

[b] Including civil engineering and test operation.

[c] Interest (10%) on initial charge of catalyst and interest and depreciation (25%) on investment cost excluding catalyst.

[d] Catalyst life: 2 years for oil and 1 year for coal.

[e] Ammonia, power, etc.

and a catalyst life of 2 years for oil and 1 year for coal, are 0.32 for oil and 0.95 for coal, while the costs per unit amount of NOx removed is just about equal for oil and coal. Compared with 80% removal, 90% removal costs about 40% more in yen/kWhr. Actually 90% NOx removal may be difficult for a large boiler without increasing leak NH_3, because gas velocity as well as NOx concentration may not be uniform in different parts of the SCR reactor inlet.

For coal, about 70% of the annualized SCR costs is accounted for by catalyst. If the catalyst is useful for 2 years, the costs will be lowered by about 35%. The catalyst life is usually guaranteed for 1 year for both oil and coal. Operation experiences have shown that the catalyst for oil may be useful for over 3 years. It may be possible to extend catalyst life for coal to 2 years.

4. SHIMONOSEKI PLANT, CHUGOKU ELECTRIC

4.1 Outline

Shimonoseki Station of the Chugoku Electric Power Co. has two boilers -- a 175 MW coal-fired boiler (No. 1) and a 400 MW oil-fired boiler (No. 2). Regulations for the station are shown in the following table.

Table 7 Regulations for Shimonoseki Station

Air pollution control	
k Value	2.7 (Ground level concentration 0.0047 ppm)
SOx (total)	Below 412 Nm^3/hr
Particulates	Below 130 kg/hr
No. 1 Boiler	Below 200 mg/Nm^3
No. 2 Boiler	Below 40 mg/Nm^3
NOx	Below 330 Nm^3/hr
No. 1 Boiler	Below 350 ppm
No. 2 Boiler	Below 170 ppm
Floating particulates	Below 0.2 mg/Nm^3

Water pollution control

pH	5.8 - 8.6
Suspended solids	Below 12 kg/day Below 15 mg/liter
Normal-hexane-soluble material	Below 0.8 kg/day Below 1 mg/liter
Chemical oxygen demand	Below 12 kg/day Below 15 mg/liter

The No. 1 boiler was completed in 1967 and was burning coal and oil in the ratio of 25 to 75 before a full scale FGD plant was completed in July 1979 using the MHI wet limestone-gypsum process. After the FGD plant was put into operation, coal and oil was used in the ratio of 50 to 50. It was difficult to use larger amounts of coal because of the NOx regulation (below 350 ppm). Although the regulation may be met by combustion modification even with the burning of coal only, it was likely that further NOx reduction might be required in future. Chugoku Electric, therefore, decided to install a full-scale SCR unit, which was completed in March 1980 to allow combustion of coal only. The SCR unit is the first full-scale plant for a coal-fired boiler in the world and has the nature of a demonstration plant.

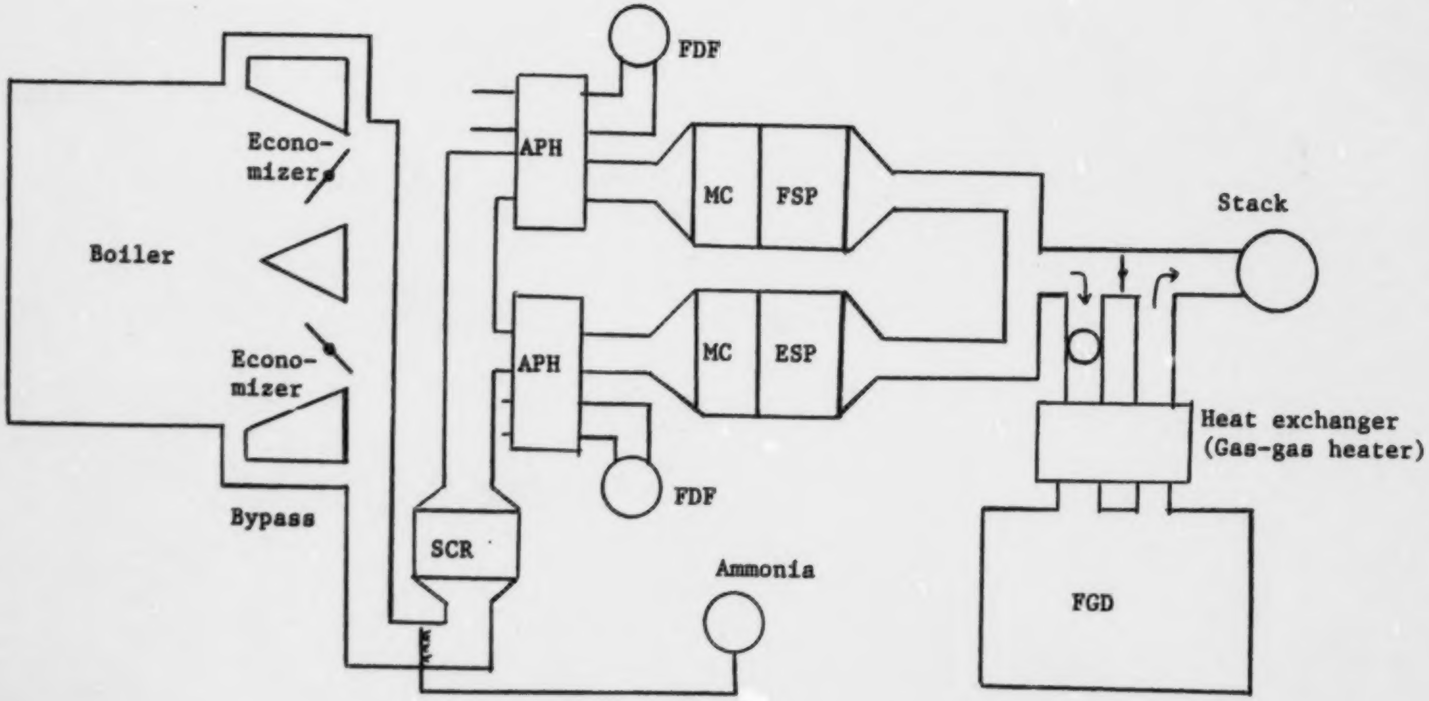
Figure 3 shows the combined system of SCR and FGD for the No. 1 boiler. The flue gas is first subjected to SCR at 330 - 400°C, passed through two trains of air preheaters and dust collectors (multicyclone and ESP), and then undergoes FGD after it is passed through a heat exchanger.

The No. 2 boiler is a relatively new one and has used a high-sulfur oil with FGD by the MHI wet limestone-gypsum process.

4.2 SCR System

The design basis of the SCR system is shown below:

Boiler capacity	175 MW
Fuel	Coal
Gas flow rate	550,000 Nm^3/hr
Gas temperature	370°C
Inlet NOx	500 ppm
Outlet NOx	250 ppm (100 ppm in future)


FDF
Econo-
mizer
APH
MC
FSP
Stack
Boiler
APH
MC
ESP
Econo-
mizer
Heat exchanger
(Gas-gas heater)
FDF
Bypass
SCR
Ammonia
FGD

SCR : Selective catalytic reduction of NOx   MC : Multicyclone

Figure 3 Flue gas treatment system for No.1 coal-fired boiler (175 MW)
(Shimonoseki Power Station, Chugoku Electric)

NOx removal efficiency	50% (80% in future)
Inlet SOx	1,600 ppm
Reactor	One reactor, with downflow of gas
Catalyst	Honeycomb. Square type with 10 mm pitch (about 8.2 mm opening)
Space velocity	3,000 hr^{-1}

The No. 1 boiler is for base load and the gas temperature at economizer outlet is normally around 360°C, suitable for SCR. The load is occasionally lowered to 25% of full load, resulting in the drop of the gas temperature to 300°C. Since ammonium bisulfate may deposit on the SCR catalyst during the low-load operation, a bypass system was installed as shown in Figure 3 to control the gas flow by dampers to mix a portion of hot gas with the economizer outlet gas to maintain the gas temperature.

An SCR reactor was installed beside the boiler so that the treated gas is sent to the existing air preheaters. The reactor contains 5 horizontal layers of honeycomb catalyst, through which flue gas is passed downwards. The flue gas contains about 410 ppm NOx, 360 ppm SO_4 and nearly 20 grams/Nm^3 of fly ash. A layer of "dummy" spacer with the same shape as the honeycomb was placed on top of the first honeycomb layer, in order to maintain a uniform parallel gas flow and to prevent catalyst erosion by fly ash.

Planning and design of the SCR system was started in July 1979. Construction was begun in October 1979. Boiler modification and reactor connection were performed during the shutdown of the boiler for annual maintenance between February 1 and March 31, 1980. Since start-up of operation in April 1980, the boiler, the SCR system and the FGD system have been operated without trouble.

Current regulations require about 50% NOx removal. Therefore, a NH_3/NOx mole ratio of about 0.56 has been used to reduce NOx concentrations from 410 to 185 ppm (55% removal) and to maintain leak NH_3 at reactor outlet below 3 ppm. In future, 80% of NOx may be removed by increasing the amount of catalyst and by using about 0.82 mol NH_3 to 1 mol NOx, keeping leak NH_3 below 5 ppm.

A catalyst life of 1 year is guaranteed by MHI, which will take all of the spent catalyst when fresh catalyst is placed. Replacement of catalyst will require 15 days with 15 workers working 7 hours a day.

The air preheater has had a soot blow system on the cold side which has been operated 4 times a day, two hours each time. When the SCR system was installed, an additional soot blow system was installed on the hot side of the preheater, which has also been operated 4 times a day, 2 hours each time. The plugging problem of the preheater by ammonium bisulfate has thus been prevented. The soot blow system will be used less frequently.

The total investment cost was about 2 billion yen including the boiler modification of which 1.7 billion was paid to the constructor.

4.3 FGD System

A flow sheet of the FGD system is shown in Figure 4. Flue gas leaving the air preheater at 160°C is cooled to about 95°C by a Ljungstrom type heat exchanger and introduced into a semiventuri type spray scrubber newly developed by MHI for particulate removal, and then into a grid packed tower with a holding tank at the bottom and a mist eliminator at the top. Limestone slurry is fed to the tank. The treated gas at 55°C is heated to 120°C by the heat exchanger eliminating gas heating by oil firing. About 90% of both SO_2 and particulates are removed (Tables 3 and 5). Slurry handling systems -- oxidation of calcium sulfite, gypsum centrifuge, etc., are similar to those of the standard MHI process.2)

After its startup in July 1979, the FGD plant was operated continuously without trouble until February 1980, when the boiler was shut down for annual maintenance. During the operation period, coal and oil were used in the ratio of 25 to 75 at the beginning and then in the ratio of 50 to 50. Fresh water, at the rate of 30 tons/hr, was fed mainly to the syray tower and used partly for mist eliminator wash. Of the 30 tons/hr, 13 tons were volatilized, 2 tons went into gypsum as water of crystallization and moisture, and 15 tons were sent to a wastewater treatment system.

Inspection during the shutdown period detected a little deposit of particulates in the heat exchanger and a slight erosion of rubber lining but neither scaling nor corrosion. The soot blow system was reinforced during the shutdown period in order to eliminate the deposit formation in the heat exchanger.

Since its restart in April, using coal only this time, the FGD system has been operated trouble-free again. Because a fan is placed upstream of the heat exchanger, a small amount of inlet gas at 160°C leaks in the heat exchanger to mix with the FGD outlet gas, thus lowering the removal efficiency of SO_2 and particulates to some extent (Table 8). Placing the fan between the heat exchanger and the prescrubber (cooler) results in the leak of the FGD outlet gas to the inlet and an increase in removal efficiency, but it may cause corrosion of the fan due to condensation of sulfuric acid at low temperatures around 90°C. MHI has been testing a new type of air preheater without gas leakage.

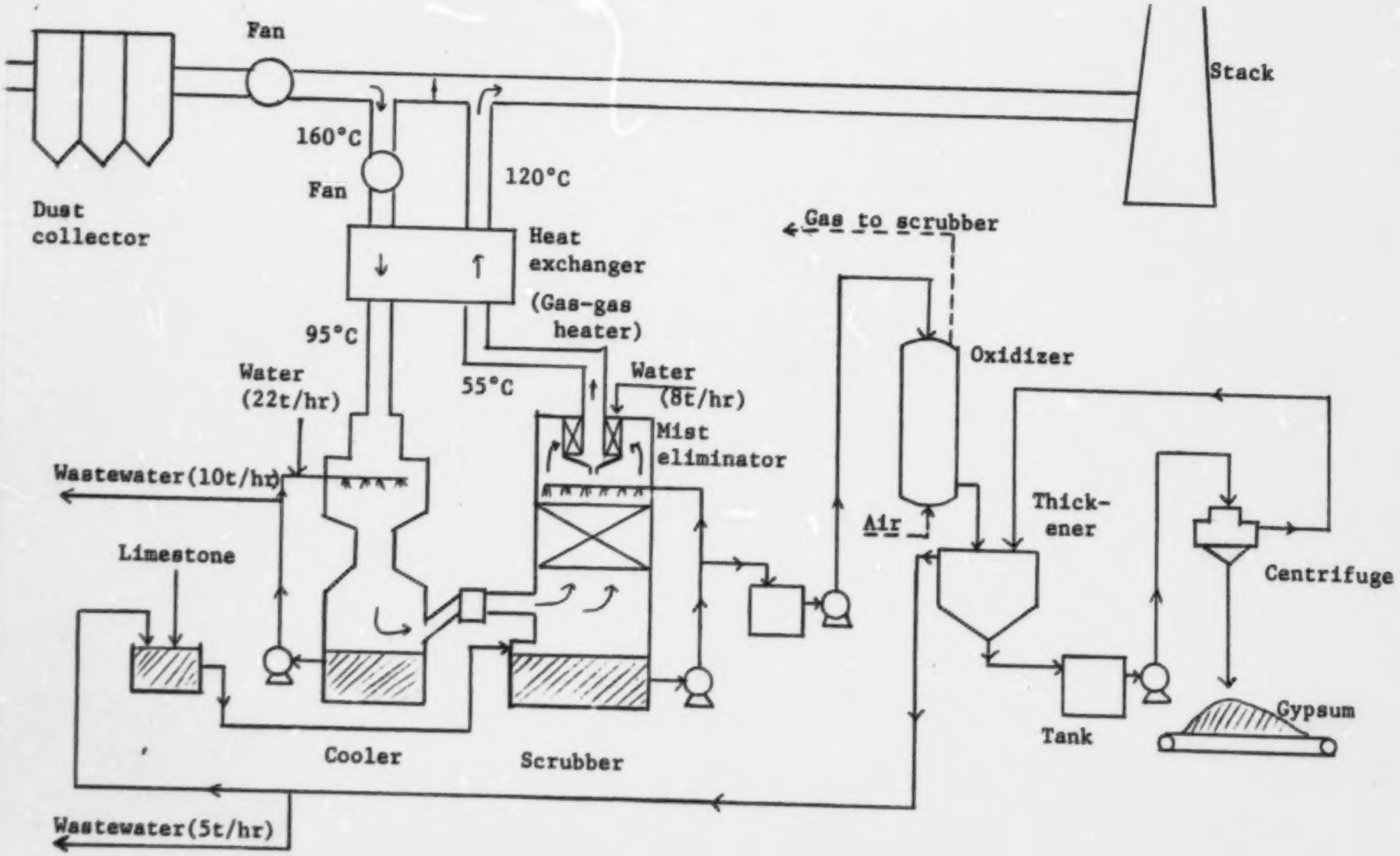

Fan
Stack
160°C
Fan
120°C
Dust collector
Heat exchanger
(Gas-gas heater)
Gas to scrubber
95°C
Oxidizer
Water (22t/hr)
55°C
Water (8t/hr)
Mist eliminator
Wastewater(10t/hr)
Thick-ener
Air
Limestone
Centrifuge
Tank
Gypsum
Cooler
Scrubber
Wastewater(5t/hr)

Figure 4 Flowsheet of FGD system for No.1 boiler at Shimonoseki Power Station

Table 8 SO_2 and particulate removal efficiency(Shimonoseki plant)

Pollutants	Concentration and removal efficiency	Coal and oil (50 : 50)	Coal only Low S	Coal only Medium S
SO_2	FGD inlet (ppm)	1,230	355	1,310
	FGD outlet (ppm)	78	20	55
	Removal efficiency(%)	93.7	94.4	95.8
	HE[a] outlet (ppm)	136	38	115
	Removal efficiency(%)	89.0	89.2	91.2
Particulates	FGD inlet (mg/Nm^3)	200	1280	830
	FGD outlet(mg/Nm^3)	12	80	50
	Removal efficiency(%)	94.0	93.8	94.0
	HE[a] outlet (mg/Nm^3)	21	130	85
	Removal efficiency(%)	89.5	89.8	89.7

---

a Heat exchanger

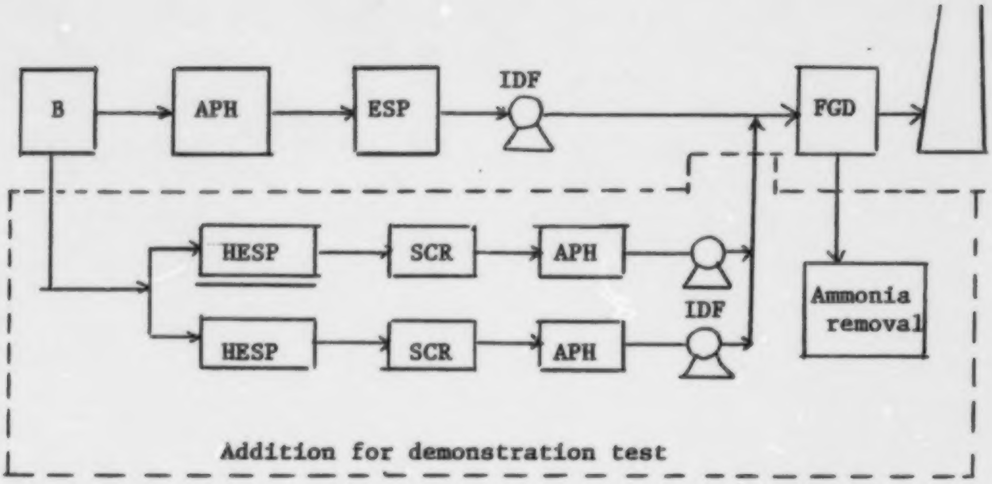
Ammonia contained in a small amount in flue gas has had no adverse effects on FGD and on the quality of fly ash which has been used for cement and land fill. Also, ammonia has been injected into the flue gas from the No.2 oil-fired boiler between the air preheater and ESP in order to prevent corrosion of ESP and to increase soot removal efficiency. Thus ammonia is contained in the flue gas introduced into the No.2 FGD system, which has also been operated without trouble.

Chugoku Electric recently decided to install similar SCR and FGD system for 5 relatively small existing coal-fired boilers.

5. OTHER COMBINED SYSTEMS

5.1 Takehara Plant, EPDC

EPDC has been constructing a full-scale demonstration plant of SCR combined with FGD at its Takehara Station for the No. 1 boiler (250 MW). Since various types of coals including low-sulfur coal will be used, a hot electrostatic precipitator is installed. As shown in Figure 5, all of the flue gas from the boiler is passed through two parallel trains of a hot ESP, SCR reactor, air preheater and ID fan. One of the reactors is constructed by Babcock Hitachi Ltd. using a plate catalyst developed by Hitachi Ltd., while the other is constructed by Kawasaki Heavy Industries (KHI) using a tubular catalyst. Over 80%


B
APH
ESP
IDF
FGD
HESP
SCR
APH
HESP
SCR
APH
IDF
Ammonia removal
Addition for demonstration test

B Boiler, APH Air preheater, ESP Electrostatic precipitator
IDF Induced fan, HESP Hot electrostatic precipitator

Figure 5 Demonstration plant at Takehara, EPDC (250 MW)

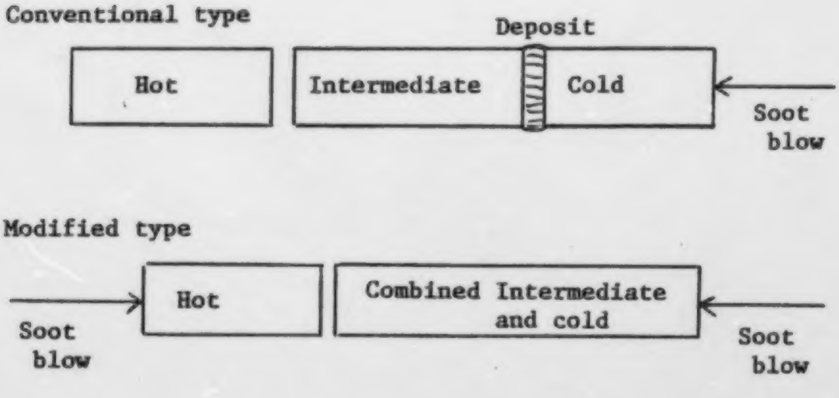

Conventional type
Deposit
Hot
Intermediate
Cold
Soot blow
Modified type
Hot
Combined Intermediate and cold
Soot blow
Soot blow

Figure 6 Arrangement of air preheater elements

of NOx will be removed maintaining leak NH_3 below 10 ppm.

Since the air preheater treats an SOx-rich, dust-lean gas, ammonium bisulfate may deposit in intermediate and low temperature zones (Figure 6). Pilot plant tests have shown that the deposit formed between the two zones is difficult to remove by soot blowing. For the demonstration plant, a modified design of the air preheater elements as shown in Figure 6 will be used to reduce the plugging problem.

The treated gas is sent to an existing FGD plant constructed by Babcock Hitachi using the limestone-gypsum process (Table 3). The leak NH_3 will be caught by the FGD system and contained in the wastewater. EPDC has installed a wastewater treatment system using a conventional activated sludge process to remove ammonia, because Takahara Station faces the Seto Inland Sea which is sometimes plagued by the red tide problem.

The total additional system for the demonstration as shown in Figure 5 cost 8 billion yen including control systems and a storage and injection system of ammonia. The new ID fans are estimated to consume about 1,500 kW more than does the existing ID fans, which is equivalent to 0.6% of the power generated by the boiler.

EPDC will construct a full scale combined system for the new No. 3 boiler (700 MW), for which the low-dust system may also be applied.

5.2 Tomato-Atsuma Plant, Hokkaido Electric

Hokkaido Electric Power Co. has constructed a new 350 MW coal-fired boiler in a newly opened industrial region near Tomakomai, which has started test operation in summer 1980 and is scheduled to be put in commercial operation in October 1980 using a low-sulfur coal (S = 0.3%). By an agreement with local governments, SOx emissions should be kept below 180 Nm^3/hr (about 140 ppm), NOx below 200 Nm^3/hr (about 160 ppm), and particulates below 200 kg/hr (about 160 mg/Nm^3).

For SOx abatement, half of the gas from the boiler is treated by a wet limestone-gypsum process FGD plant constructed by Babcock Hitachi. NOx is reduced below 200 ppm by combustion modification including staged combustion, flue gas recirculation, and dual-register low-NOx burners. In addition, one-fourth of the gas is treated by SCR for 80% NOx removal to meet the agreement.

Since a low-sulfur coal is used, a hot electrostatic precipitator has been installed which reduces the dust content down to 45 mg/Nm^3. One-fourth of the gas passing through the hot ESP is treated by an SCR reactor containing a plate catalyst developed by Hitachi Ltd. An economizer bypass system has been installed to maintain the gas temperature above 300°C.

Hokkaido Electric plans to install a 600 MW coal-fired boiler. If the plan is authorized, Hokkaido Electric plans to reevaluate the design including the necessity of the bypass and the use of cold vs. hot ESP.

5.3 Nakoso Plant, Joban Joint Electric Co.

Tokyo Electric Power Co., jointly with Tohoku Electric Power Co., Joban Joint Electric Co., and MHI, has carried out pilot plant tests at Nakoso Station of Joban on combined systems of SCR (high-dust and low-dust) and wet limestone-gypsum process FGD using 4,000 Nm^3/hr of flue gas from a coal-fired boiler. In 1979, the high dust system was operated for 5,000 hours while the low-dust system was operated for 4,000 hours. Further tests are in progress in 1980.

Honeycomb catalysts are used for both systems with downflow of the gas. With the high-dust system, erosion of the catalyst by dust has been prevented by placing on top of the honeycomb a dummy spacer which has the same cross section as the honeycomb. The air preheater has been kept clean by applying soot blowing once a day; ammonium bisulfate has not deposited appreciably because of the cleaning effect of fly ash. With the low-dust system, the dust leaving the hot ESP is in a small amount but consists of fine particles which are rather sticky and tend to deposit particularly at the inlet of the honeycomb. Moreover, the air preheater requires soot blowing 3 times a day to prevent the deposit of ammonium bisulfate.

The FGD system has been operated without trouble. A semiventuri type spray scrubber developed by MHI is used for the prescrubbing. Tests indicated that the dust contained in the gas in concentrations of 100, 200, and 300 mg/Nm^3 was reduced to about 20, 30, and 40 ppm, respectively, by the prescrubber and to about 15, 20, and 30 ppm, respectively by the SO_2 absorber.

Joban has started to construct 2 new boilers with a capacity of 600 MW each, which will use low-sulfur oil with a small amount of coal to start with. Both boilers will have high-dust system SCR units with a honeycomb catalyst. The units for one of the boilers will be constructed by MHI and the units for the other boiler by IHI. FGD plants may be constructed when larger amounts of coal are used.

6. OTHER MAJOR ACTIVITIES

6.1 Pilot Plant Tests by Activated Carbon Process[3)]

EPDC, jointly with Sumitomo Heavy Industries, has been operating a pilot at Takehara with a capacity of treating 10,000 Nm^3/hr of flue gas from the No. 1 coal-fired boiler to remove over 90% of SO_2 and over 30% of NOx by activated carbon and ammonia. A flowsheet of the pilot plant is shown in Figure 7. The flue gas containing 1,300 ppm of SO_2 and 320 ppm of NOx at about 150°C is injected with 225 ppm NH_3 and is introduced in a reactor with activated carbon in a moving bed. Over 90% of SO_2 is adsorbed by the carbon to form sulfuric acid and ammonium sulfate (reactions 1 and 2) while over 30% of NOx is converted to N_2 (reaction 3).

$$SO_2 + H_2O + 1/2\ O_2 \rightarrow H_2SO_4 \quad \text{............} \quad (1)$$

$$H_2SO_4 + 2NH_3 \rightarrow (NH_4)_2SO_4 \quad \text{........} \quad (2)$$

$$4NO + 4NH_3 + O_2 \rightarrow 4N_2 + 6H_2O \quad \text{.......} \quad (3)$$

The char loaded with the sulfur compounds is heated in a separate moving bed to over 350°C by inert gas produced by incomplete combustion of LPG gas. Concentrated SO_2 gas is released by the heating (reactions 4 and 5), then is introduced into a coal-bed reactor and converted to S by the Resox process developed by Foster Wheeler Co. (reaction 6). The sulfur vapor is condensed to recover elemental sulfur. The gas leaving the condenser is incinerated and sent to the existing wet limestone-gypsum process FGD plant.

$$H_2SO_4 + 1/2\ C \rightarrow SO_2 + 1/2\ CO_2 + H_2O \quad \text{..........} \quad (4)$$

$$(NH_4)_2SO_4 + O_2 \rightarrow SO_2 + N_2 + 4H_2O \quad \text{..............} \quad (5)$$

$$SO_2 + C \rightarrow S + CO_2 \quad \text{......................} \quad (6)$$

About 1.6% of the carbon is consumed in one cycle which takes 3 days. The sulfur condenser had a plugging problem, which has been solved by applying a technology used for the Claus furnace. The remaining major problem is the low recovery of sulfur at 60 - 70%. Efforts have been made to improve the recovery.

The low NOx removal efficiency is due to the low temperature. Over 200°C with over 2 mole NH_3 to 1 mol NOx may be needed to attain over 80% removal. For commercial application, it may be preferable to use SCR for the boiler economizer outlet at 300 - 400°C and then apply the carbon process for SO_2 removal only without using ammonia. EPDC

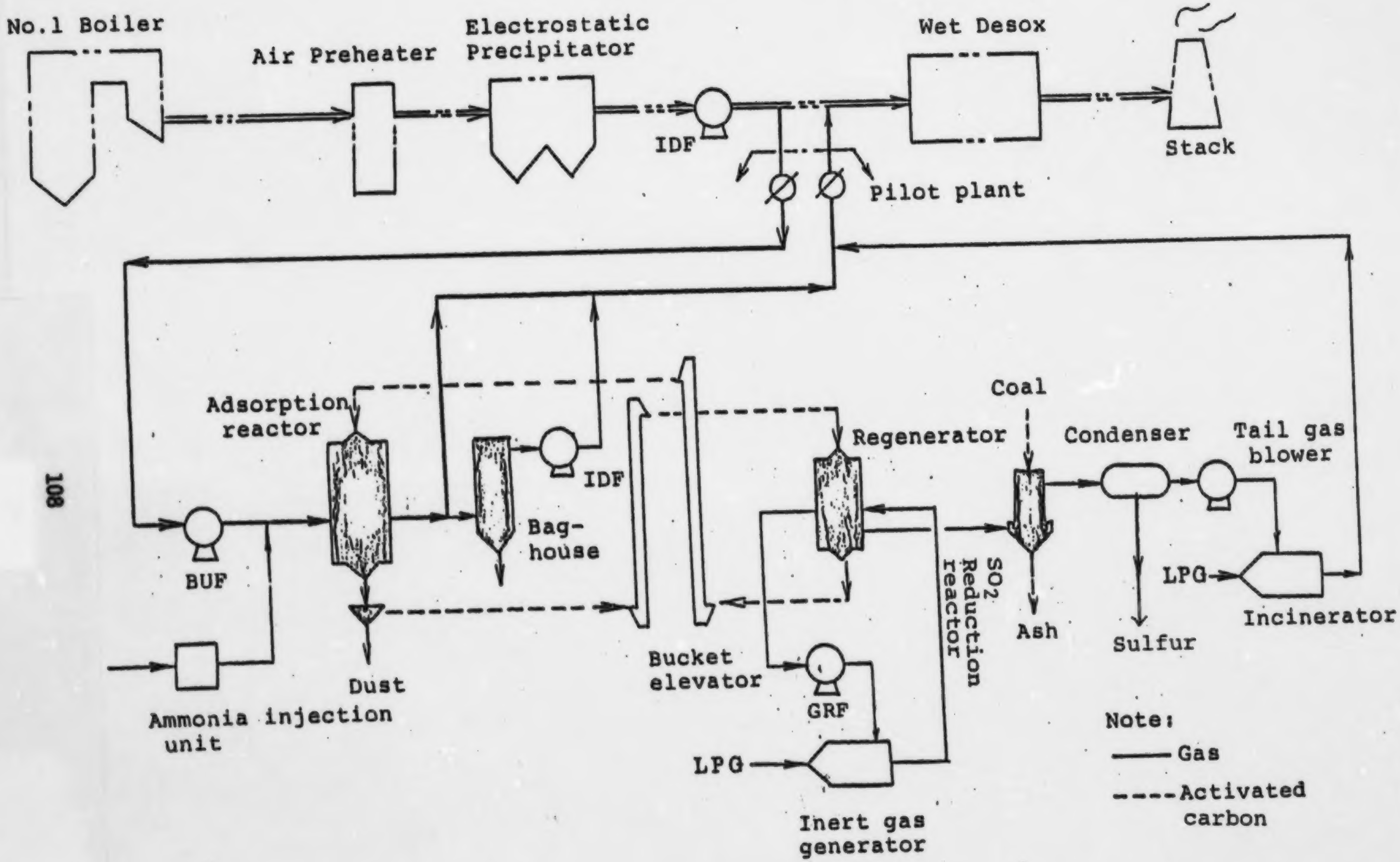

No.1 Boiler
Air Preheater
Electrostatic Precipitator
IDF
Wet Desox
Stack
Pilot plant
Adsorption reactor
BUF
Ammonia injection unit
Dust
Bag-house
IDF
Bucket elevator
Regenerator
GRF
LPG
Inert gas generator
SO_2 Reduction reactor
Coal
Ash
Condenser
Sulfur
Tail gas blower
LPG
Incinerator
Note:
Gas
Activated carbon

Figure 7 Pilot plant for FGD by activated carbon process for elemental sulfur recovery 3)
(Takehara plant, EPDC)

is to install a prototype plant of the carbon process at its Matsushima Station by 1982 to treat one-fourth of the gas from a new 500 MW coal-fired boiler, while three-fourths of the gas will be treated by the wet limestone-gypsum process.

6.2 New Combustion Technology

About one-tenth of fuel used for the boiler is injected above the combustion zone in the boiler to form a reducing atmosphere where NOx formed by the combustion is reduced to N_2. Air is added above the reducing zone for complete combustion. The technology was originated by MHI and has been further developed by Tokyo Electric Power Co. jointly with MHI, Hitachi, and IHI for NOx abatement for boilers. Tests with pilot plants with a capacity ranging from 5,000 to 8,000 kW using various fuels have indicated that about 50% of NOx is removed. By using the process in combination with conventional combustion modification, NOx concentration has been reduced to 10 - 20 ppm with gas, 40 - 60 ppm with oil, and 60 - 100 ppm with coal. The boiler is a little larger than a conventional boiler. Tests on a larger scale are planned.

REFERENCES

1. Y. Nakabayashi, Plan, Design and Operating Experience of FGD For Coal Fired Boilers Owned by EPDC, Paper No. 41, EPA FGD Symposium, March 1979

2. J. Ando, SO_2 Abatement for Stationary Sources in Japan, EPA-600/7-78-210, November 1978

3. EPDC and Sumitomo Heavy Industries,.Simultaneous SOx-NOx Removal System for Coal-fired Boiler, October 1979

Session 2: IMPACT OF RECENT LEGISLATION/REGULATIONS

Walter C. Barber, Chairman
Office of Air Quality Planning and Standards
U. S. Environmental Protection Agency
Research Triangle Park, North Carolina

Panel: Impact of Recent Legislation/Regulations
Brief overviews of recent Legislation/Regulation, under the CAA, CWA, and RCRA, followed by questions from the audience.

Members: John W. Lum
Office of Water Planning and Standards
U.S. Environmental Protection Agency
Washington, D.C.

Penelope Hansen
Office of Solid Waste
U.S. Environmental Protection Agency
Washington, D.C.

No papers or discussions are included for this session.

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Session 3: FGD RESEARCH AND DEVELOPMENT PLANS

Julian W. Jones, Chairman
Industrial Environmental Research Laboratory
U. S. Environmental Protection Agency
Research Triangle Park, North Carolina

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RECENT TRENDS IN UTILITY FLUE GAS DESULFURIZATION

by

M. P. Smith, M. T. Melia, and B. A. Laseke, Jr.
PEDCo Environmental, Inc.

and

Norman Kaplan
U.S. Environmental Protection Agency
Industrial Environmental Research Laboratory
Emissions/Effluent Technology Branch

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ABSTRACT

PEDCo Environmental, Inc., under contract to the Industrial Environmental Research Laboratory-RTP and the Division of Stationary Source Enforcement of the U.S. Environmental Protection Agency, has been monitoring the status of utility flue gas desulfurization (FGD) technology since 1974. Information for this program is obtained by visits to plants having operational FGD systems and through periodic contacts with representatives of utility companies, FGD system and equipment suppliers, system designers, research organizations, and regulatory agencies.

This paper summarizes the status of utility FGD technology as of the end of August 1980 and indicates recent trends in both the design and performance of the FGD systems. The discussion of current status includes the number and capacity of operational and planned FGD systems, as well as identification of the systems according to process type, emission control strategy, SO_2 inlet concentration levels, and removal efficiencies. Process design developments and trends are summarized for the major components and subsystems associated with commercial FGD systems. In discussing FGD system performance, composite graphs are included presenting annual system availability data (through June 1980) for low-, medium-, and high-sulfur coal FGD installations. A statistical analysis of the data for the years 1978 and 1980 indicates overall trends in FGD system dependability. Finally, capital and annual cost data (both reported and adjusted) are included for the operational FGD systems and cost model comparisons are made.

The current data indicate that 203 FGD systems are either operational, under construction, or planned (as of August 1980), representing a total controlled capacity of about 97,000 MW. Of the 203, 73 systems are operational, representing 27,155 MW of controlled capacity. The dependability analysis indicates that the overall median availability for these operational systems has increased 1.5%, 16.5%, and 50.6% for low-, medium-, and high-sulfur coal FGD installations, respectively, between the years 1978 and 1980.

NOTES

1. Company Names and Products.

   The mention of company names or products is not to be considered an endorsement or recommendation for use by the U.S. Environmental Protection Agency.

2. Consistency of Information.

   The information presented was obtained from a variety of sources (sometimes by telephone conversation) including system vendors, users, EPA trip reports and other technical reports. As such, consistency of information on a particular system and between the several systems discussed may be lacking. The information presented is basically that which was voluntarily submitted by developers and users with some interpretation by the author. The order of presentation of information or the amount of information presented for any one system should not be construed to favor or disfavor that particular system.

3. Units of Measure.

   EPA policy is to express all measurements in Agency documents in metric units. When implementing this practice will result in undue cost or difficulty in clarity, IERL-RTP provides conversion factors for the non-metric units. Generally, this paper uses British units of measure.

   The following equivalents can be used for conversion to the Metric system:

British	Metric
5/9 (°F-32)	°C
1 ft	0.3048 m
1 ft^2	0.0929 m^2
1 ft^3	0.0283 m^3
1 grain	0.0648 gram
1 lb (avoir.)	0.4536 kg
1 ton (long)	1.0160 m tons
1 ton (short)	0.9072 m tons
1 gal.	3.7853 liters
1 lb/10^6 Btu	429.6 ng/J
1 Btu/kWh	1055.056 J/kWh

SECTION 1

INTRODUCTION

For more than 6 years PEDCo Environmental, Inc., under contract to the U.S. Environmental Protection Agency (EPA), has monitored the development and growth of flue gas desulfurization (FGD) technology for fossil fuel-fired utility boilers in the United States. The program provides an objective and current perspective of FGD technology as applied to fossil fuel-fired utility boilers and facilitates, through information dissemination, improvements in the design and performance of current and future systems.

The program addresses performance of operational FGD systems, process and design characteristics of both operational and planned systems, projected application and nature of future processes and systems, and costs associated with both current and planned systems. The program also includes the monitoring of particulate matter scrubbers operating on coal-fired utility boilers in the United States and FGD systems operating on coal-fired utility boilers in Japan.

Program emphasis is on the performance of the operational systems. Accurate portrayal of system performance requires data concerning system/module dependability, operating problems and solutions, operating and maintenance costs, and outlet emissions and removal efficiency. Data on outlet emissions of sulfur dioxide (SO_2), particulate matter, and nitrogen oxides (NO_x) and on removal efficiency of SO_2 and particulate matter are considered information needs in order to assess actual system performance with respect to control requirements in the recently promulgated revised New Source Performance Standards (NSPS) for electric utility steam generating units.

Utilities, system and equipment suppliers, system designers, research organizations, regulatory agencies, and others all volunteer the information for this program. This voluntary approach facilitates timely dissemination of pertinent information in this key technological area. All information that is gathered is stored in a computerized data base known as the Flue Gas Desulfurization Information System (FGDIS). This system is discussed in more detail in Appendix A.

Information on operational systems is verified solely by the utilities and reported essentially as received. Any modifications or adjustments to the reported data are made solely for purposes of a consistent format that will allow reliable comparisons and evaluations to be made.

SECTION 2

TECHNOLOGY OVERVIEW

CURRENT STATUS

Table 2-1 lists the number of domestic utility FGD systems according to status and equivalent electrical capacities as of the end of August 1980.

TABLE 2-1. NUMBER AND TOTAL CAPACITY OF FGD SYSTEMS, AUGUST 1980

Status	No. of units	Total controlled capacity, MW[a]	Equivalent scrubbed capacity, MW[b]
Operational	73	27,155	24,765
Under construction	39	17,855	16,854
Planned:			
Contract awarded	29	13,769	12,919
Letter of intent	7	5,590	5,590
Requesting/evaluating bids	15	8,424	8,424
Considering only FGD	40	24,200	23,980
TOTAL	203	96,993	92,532

[a] Total controlled capacity (TCC) represents the gross capacities (MW) of coal-fired units brought into compliance by FGD systems, regardless of the percent of the flue gas treated.

[b] Equivalent scrubbed capacity (ESC) represents the effective capacities of the FGD systems (in equivalent MW), based on the percent of the flue gas treated.

GROWTH TRENDS

Power-Generating and FGD Capacity

As indicated in Table 2-1, 73 coal-fired power-generating units currently equipped with operational FGD systems represent a total controlled capacity of 27,155 MW. This compares with a December 1979 total coal-fired power-generating capacity of approximately 235,000 MW. Current projections indicate that the latter will rise to approximately 370,000 MW by the end of 1990. Based on the known utility commitments to FGD, the percentage of coal-fired capacity controlled by FGD will increase from its current level of 11.5% to 26.5% by the end of 1990.

Table 2-2 presents the projected distribution of power-generating sources (by energy source) in the electric utility industry. Table 2-3 presents the percentage of current and projected coal-fired and total power-generating capacities controlled by FGD.

Based on the requirements of the revised NSPS, actual FGD-controlled capacity should exceed the levels indicated in the preceding discussion. Currently, about 50 additional units, representing a total capacity of approximately 25,000 MW, have been identified as requiring SO_2 controls in the decade just begun; however, identification of these units and information regarding their status is not ready for public release as a result of the premature stage of their planning, developments in ongoing litigation, and the determination of applicable emission control standards.

Figure 2-1 shows current and projected FGD-controlled capacity and total power-generating capacity of coal-fired units through 1990. This figure represents the committed FGD-controlled capacity (those systems identified in Table 2-1), the uncommitted FGD-controlled capacity (those units that cannot be identified at the present time), and current and projected coal-fired power generating capacities (those values cited in Table 2-2 and the preceding discussion).

Figure 2-2 shows estimated FGD-controlled capacities at the indicated month and year. An estimated total of 37,834 MW of FGD-controlled capacity was identified in November 1974. By August 1980, this figure had risen to 96,993 MW (see Table 2-1). This represents an overall growth rate of 156% for the 6-year period. In addition, the figures reflect a better than 55% increase in the last 2 years.

Other notable changes that occurred during the 1974 to 1980 growth period include:

- A 384% increase in the number of operational systems.
- A 753% increase in operating capacity (ESC).
- An increase in the average capacity of the FGD-equipped unit from 170 MW to 340 MW.

TABLE 2-2. DISTRIBUTION OF POWER-GENERATING SOURCES BY ENERGY SOURCE[a,b]

	Percent of total						Total, GW
	Coal	Nuclear	Oil	Hydro	Gas	Other	
December 1979	39	9	25	13	13	1	603
December 1990	44	14	20	11	10	1	833

[a] Adapted from U.S. Department of Energy (1979) and Rittenhouse (1978).[1,2]

[b] Figures reflect annual losses of 0.4% of the year-end capacity attributed to retirement of older units.

TABLE 2-3. FGD-CONTROLLED POWER-GENERATION CAPACITY (percent of total)

Period	Coal-fired capacity	Total capacity
August 1980[a]	11.5[b]	4.5[b]
December 1990	26.5	11.6

[a] Represents FGD-committed capacity as of August 1980.

[b] Based on FGD capacity as of August 1980 and total power-generating capacity as of December 1979.

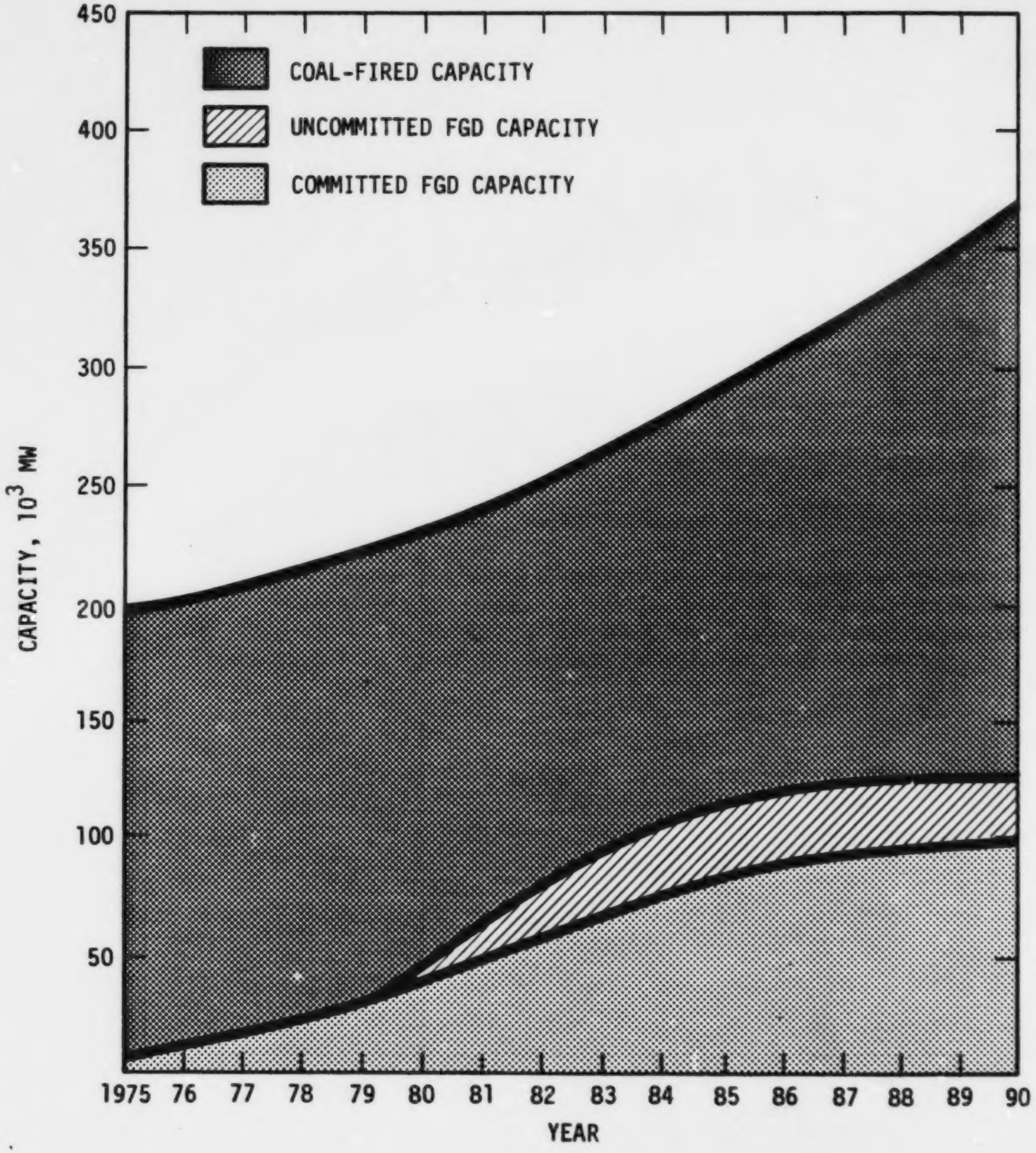

Figure 2-1. Projections of coal-fired generating capacity and FGD capacity from 1975 to 1990.

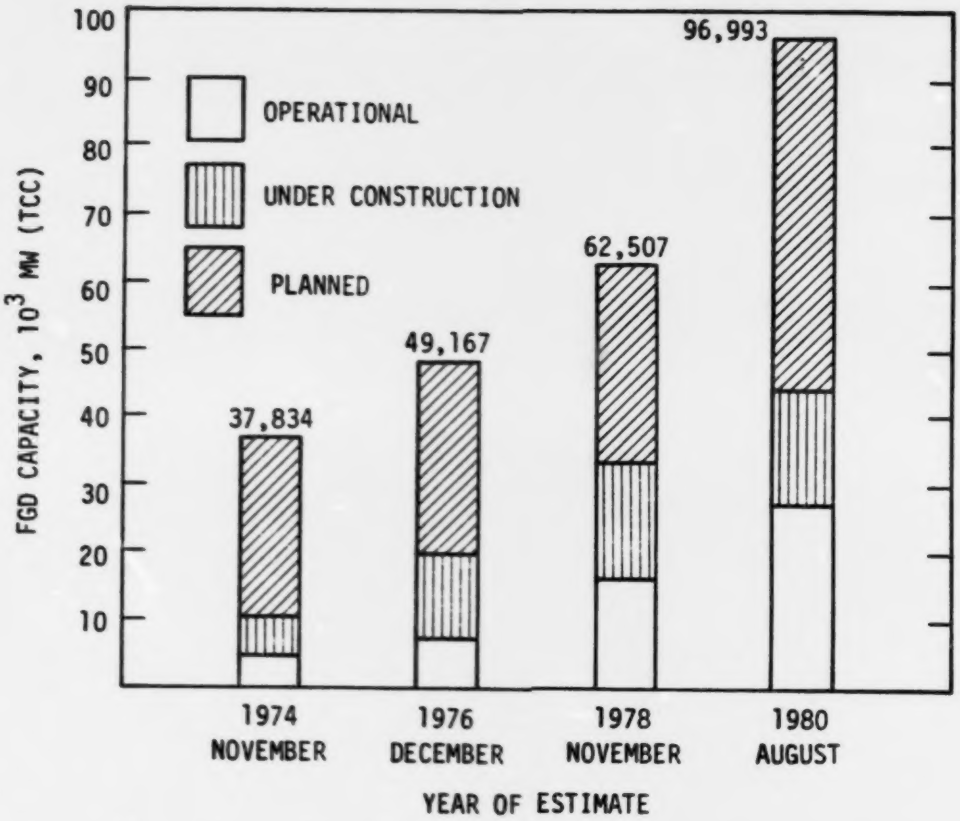

Figure 2-2. FGD capacity as a function of status and year of estimate.

Process Type

FGD systems may be categorized in several ways, some general and others more specific. Some general categorizations used in the survey are:

- wet vs. dry process
- throwaway product vs. salable product process

A more specific categorization is by process (e.g., lime, limestone, magnesium oxide, Wellman-Lord).

Tables 2-4, 2-5, and 2-6 summarize the current status of FGD capacities associated with each of the foregoing process categories. These tables show that the vast majority of operating experience to date has been obtained with wet calcium-based, throwaway-product FGD systems. Of the 68,044 MW of FGD capacity committed to a specific process (see Table 2-6), 62,541 MW (approximately 92%) are wet calcium-based, throwaway-product systems.

Table 2-4 shows that all currently operating processes are wet systems. With the recent advent of spray dryer collection processes, 10 systems, representing an ESC of 3,523 MW, are currently committed for future operation with a dry system. Therefore, dry systems represent almost 12% of the FGD capacity in the under construction and contract awarded status categories.

Table 2-5 indicates that approximately 6% of the current operating FGD-controlled capacity produces a salable product (elemental sulfur or sulfuric acid). This level of application of salable product processes is expected to remain relatively unchanged in the near future, as reflected by the 7% and 9% levels currently committed in the under construction and planned status categories. In the planned category, if the 641 MW scheduled to produce gypsum for sale are not considered (gypsum may have to be thrown away if a market is not available), the 9% is reduced to 7%.

Table 2-6 reflects several trends in the industry with respect to chemical process selection. Direct lime and limestone systems currently account for approximately 89% of the chemical processes selected, and a comparison of the two shows a distinct industry preference for the latter, which will get stronger in the near future as more systems are placed in service. This trend is evident in that 53% of the lime/limestone capacity in operation, 59% of the lime/limestone capacity under construction, and 66% of the planned lime/limestone capacity are limestone systems.*

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* Includes alkaline fly ash lime/limestone processes.

TABLE 2-4. COMMITTED FGD CAPACITY - WET VS. DRY PROCESSES

	FGD capacity (ESC), MW			
	Operational	Under construction	Contract awarded	Total
Wet	24,767	15,194	11,056	51,017
Dry	0	1,660	1,863	3,523
TOTAL	24,767	16,854	12,919	54,540

TABLE 2-5. DISTRIBUTION OF FGD SYSTEMS BY END-PRODUCT

	FGD capacity (ESC), MW			
	Operational	Under construction	Planned	Total
Salable product	1,600	1,208	2,991[a]	5,799
Throwaway product	23,167	15,646	29,678	68,491
TOTAL	24,767	16,854	32,669[b]	74,290[b]

[a] This total contains 641 MW of capacity which will produce gypsum for sale rather than sulfur or sulfuric acid.

[b] This total is less than that reflected in Table 2-1 because a number of planned FGD systems have not yet been committed to a process.

TABLE 2-6. DISTRIBUTION OF FGD SYSTEMS BY PROCESS

	FGD capacity (ESC), MW			
Process	Operational	Under construction	Planned	Total
Limestone[a]	11,172	8,816	16,164	36,152
Lime[b]	9,869	4,940	6,035	20,844
Lime/spray drying	0	1,120	1,907	3,027
Lime/limestone	20	0	475	495
Sodium carbonate	925	330	250	1,505
Magnesium oxide	0	574	750	1,324
Wellman Lord	1,540	534	0	2,074
Dual alkali	1,181	0	842	2,023
Aqueous carbonate/ spray drying[c]	0	540	0	540
Citrate[d]	60	0	0	60
Total	24,767	16,854	26,423[e]	68,044

[a] Includes alkaline fly ash/limestone and limestone slurry process design configurations.

[b] Includes alkaline fly ash/lime and lime slurry process design configurations.

[c] Includes nonregenerable dry collection and regenerable process design configurations.

[d] This system is operating at the St. Joseph Zinc Co., G. F. Wheaton Plant and is listed as a utility FGD system because the plant is connected by a 25-MW interchange to the Duquesne Light Company.

[e] Because the processes of all planned systems are not known, the totals in this status category are less than those in Table 2-1.

Emission Control Strategy

Emission control strategy refers to the measures used to control particulate matter and SO_2 emissions from power plants firing fossil fuels. At FGD-equipped, coal-fired utility boilers, three basic combinations of primary particulate matter/SO_2 control equipment are used: electrostatic precipitator (ESP)/FGD, fabric filter (FF)/FGD, and two-stage scrubbing. Table 2-7 summarizes emission control strategies for the current and planned FGD-equipped units.

TABLE 2-7. SUMMARY OF EMISSION CONTROL SELECTION[a]

	Operational		Under construction		Contract awarded		Total	
	No.	MW	No.	MW	No.	MW	No.	MW
ESP/FGD	46	16,564	32	13,890	22	10,823	100	41,277
FF/FGD			3	990	7	2,096	10	3,086
Two-stage scrubbing	27	8,203	4	1,974	0	0	31	10,177
Total	73	24,767	39	16,854	29	12,919	141	54,540

[a] Capacities represent ESC.

As indicated in Table 2-7, several industry preferences emerge with respect to selection of a control strategy. The most obvious is the strong preference to use an ESP for primary particulate matter control upstream of the FGD system. Second, a small but increasing preference for FF's is influenced by the advent of the spray dryer/dry collection FGD technology. The suppliers of most of the dry processes offered commercially recommend a FF as the preferred collection device. All the FF/FGD combinations presented in this table are spray dryer/dry collection systems. Third, a preference for the use of two-stage scrubbing system for SO_2 and particulate matter control is diminishing. The units under construction that will use two-stage scrubbing are either retrofit applications where the existing particulate matter control devices (ESP's) need upgrading or new applications where the alkalinity of the collected fly ash will be used as a source of reagent.

APPLICATION CONSIDERATIONS

New/Retrofit Units

Figure 2-3 shows a comparison of FGD-controlled capacities with new and retrofit FGD systems. As indicated in this figure, many of the original FGD systems were retrofits (e.g., retrofits accounted for 62% of the operating capacity in service in 1975). As of August 1980, new systems accounted for 75% of the operating capacity. This trend toward application of FGD systems on new sources is a result of the NSPS promulgated, pursuant to the Clean Air Act Amendments. By 1990, FGD systems installed on new boilers are expected to comprise 86% of the total.

Design SO_2 Removal, Coal Sulfur Content, and Inlet SO_2 Level

Tables 2-8 and 2-9 summarize the FGD systems in service, under construction, and planned according to design values for SO_2 removal, coal sulfur content, and inlet SO_2 level. Table 2-8 presents a breakdown of the FGD systems that are operational, under construction, and planned (contract awarded) according to level of SO_2 removal efficiency versus coal sulfur content. Some general statistics from the table are evident. First, more than 70% of the FGD capacity is designed for SO_2 removal efficiencies of 80% or greater (almost evenly distributed between efficiencies of 80 to 89% and the 90% or greater). Second, more than 85% of the FGD capacity installed or planned is for boilers burning low- and high-sulfur coals, with the capacities almost equally distributed between the two.

Table 2-9 presents a breakdown of FGD capacity by status category according to design inlet SO_2 levels. Establishing 4 $lb/10^6$ Btu as the break-off level between low- and high-inlet SO_2 leads to the conclusion that FGD systems are used to a greater extent on low-level SO_2 inlets than on high-level inlets. Since 56% of present operational capacity is applied to low-inlet SO_2 levels, as are 62% of the systems under construction, and 64% of the planned systems, it appears that more of the future coal fired utility units are expected to use low- or medium-sulfur coal with FGD than high-sulfur coal and FGD. This may be because there is more coal-fired utility growth where low- or medium-sulfur coal exists.

Note that the preferences and trends cited in Tables 2-8 and 2-9 virtually exclude any impact that may be brought about by the revised NSPS of June 1979. This discussion is therefore limited to technological preferences and trends that developed largely in response the Federal, state, and local regulatory standards under the original NSPS of December 1971.

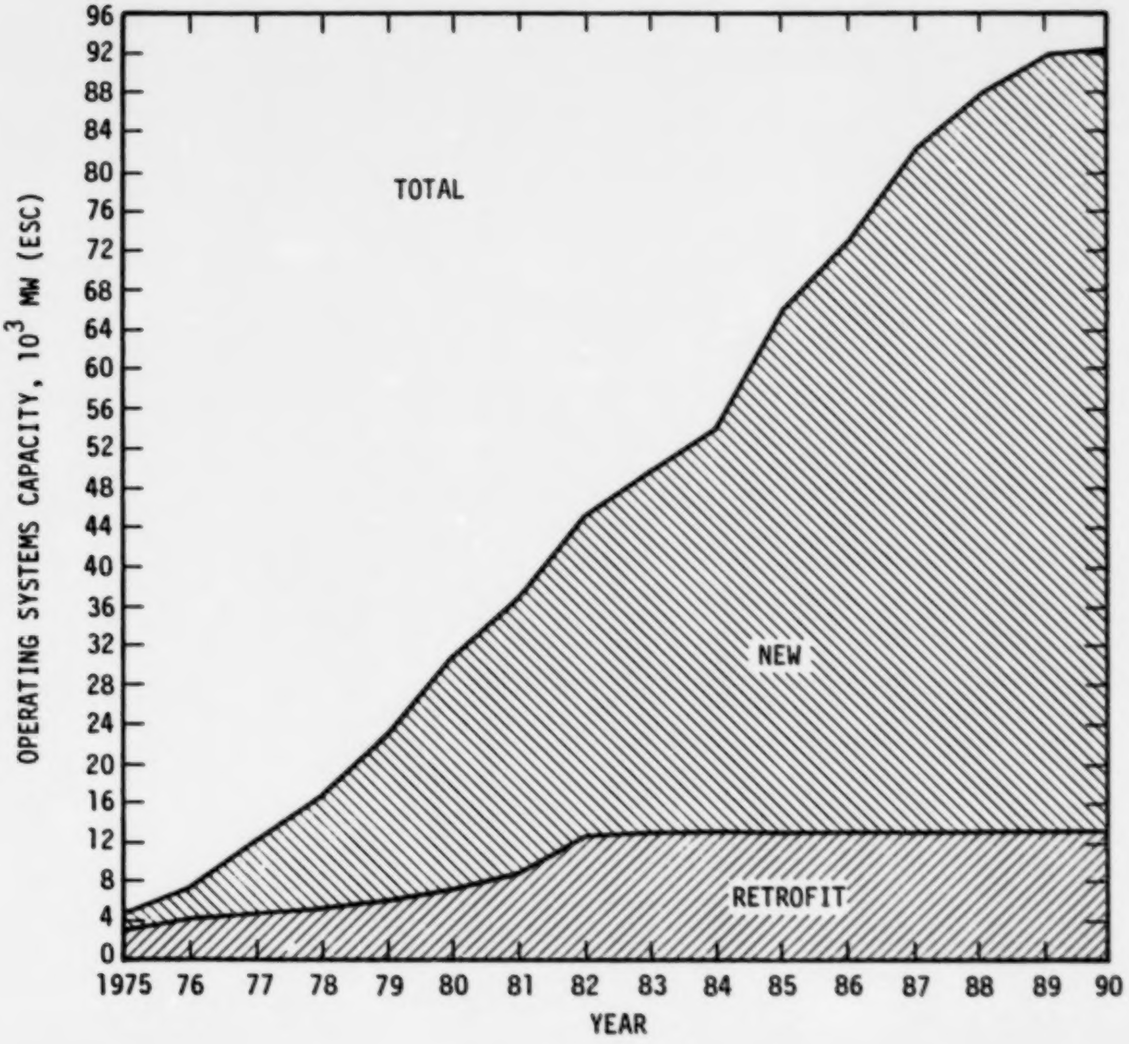

Figure 2-3. Committed FGD operating capacity for new and retrofit installations through 1990.

TABLE 2-8. DESIGN SO_2 REMOVAL EFFICIENCIES OF FGD SYSTEMS WITH RESPECT TO COAL SULFUR CONTENT

Design removal efficiency	Coal sulfur content[a]	Operational		Under construction		Contract awarded		Total	
		No.	MW[b]	No.	MW[b]	No.	MW[b]	No.	MW[b]
< 70	Low	7	3,066	2	767	0	0	9	3,833
	Medium	7	1,306	1	280	0	0	8	1,586
	High	0	0	0	0	0	0	0	0
Total		14	4,372	3	1,047	0	0	17	5,419
70-79	Low	6	2,359	3	1,262	7	3,273	16	6,894
	Medium	1	800	1	382	0	0	2	1,182
	High	4	1,180	1	500	0	0	5	1,680
Total		11	4,339	5	2,144	7	3,273	23	9,756
80-89	Low	13	3,938	2	1,017	6	3,303	21	8,258
	Medium	2	918	3	1,080	2	1,000	7	2,998
	High	12	4,181	8	3,557	4	1,955	24	9,693
Total		27	9,037	13	5,654	12	6,258	52	20,949
≥ 90	Low	6	2,044	6	3,200	2	800	14	6,044
	Medium	3	749	3	544	2	530	8	1,823
	High	12	4,225	9	4,265	6	2,058	27	10,548
Total		21	7,018	18	8,009	10	3,388	49	18,415
TOTAL	Low	32	11,407	13	6,246	15	7,376	60	25,029
	Medium	13	3,773	8	2,286	4	1,530	25	7,589
	High	28	9,586	18	8,322	10	4,013	56	21,921

[a] Low-sulfur content is less than 1%; medium-sulfur content is 1 to 2.5% sulfur; high-sulfur content is greater than 2.5%.

[b] Capacities represent ESC.

TABLE 2-9. FGD SYSTEM SO_2 INLET LEVELS

FGD system SO_2 inlet ($lb/10^6$ Btu)	Operational		Under construction		Contract awarded		Total	
	No.	MW[a]	No.	MW[a]	No.	MW[a]	No.	MW[a]
≤ 1.9	26	8,636	10	5,039	12	5,856	48	19,531
2.0 - 3.9	18	5,235	10	2,933	5	2,520	33	10,688
4.0 - 5.9	8	4,260	4	1,204	10	3,743	22	9,207
≥ 6.0	21	6,635	15	7,678	2	800	38	15,113
TOTAL	73	24,766	39	16,854	29	12,919	141	54,539

[a] Capacities represent ESC.

SECTION 3

PROCESS DESIGN DEVELOPMENTS

This section addresses preferences and trends in the process design development of commercial FGD systems.

CHEMICAL ADDITIVES

Chemical additives are used to improve the chemistry of lime- and limestone-based FGD systems. For example, magnesium-promoted processes have been used to reduce scaling, to increase sulfur dioxide removal, and to improve reagent utilization.

Table 3-1 lists the number and generating capacity of units that now have or will have FGD systems with magnesium-promoted processes.

TABLE 3-1. NUMBER AND CAPACITY OF UNITS USING MAGNESIUM-PROMOTED FGD PROCESSES

Process	Operational		Under construction		Contract awarded	
	No.	MW[a]	No.	MW[a]	No.	MW[a]
Lime	7	4,433	2	860	0	0
Limestone	0	0	1	670	1	650
Lime/alkaline fly ash	0	0	2	1,400	0	0
Total	7	4,433	5	2,930	1	650

[a] Equivalent scrubbed capacity.

The introduction of magnesium into lime- and limestone-based FGD processes has been of great interest over the last 10 years, but most full-scale magnesium-promoted systems actually began operations in the mid to late 1970's. Table 3-1 shows that the trend in the use of magnesium promotion is declining.

SYSTEM ENERGY CONSUMPTION

Table 3-2 shows the range and average of energy requirements of lime and limestone processes as a percentage of gross generating capacity for new and retrofit systems. As shown in the table, there is no significant difference between new and retrofit systems.

TABLE 3-2. ENERGY CONSUMPTION FOR OPERATIONAL WET LIME AND LIMESTONE SCRUBBING SYSTEMS[a]

Process	New[b]		Retrofit[b]		Overall[b] Average
	Range	Average	Range	Average	
Lime	1.6 - 6.0	3.8	1.5 - 3.5	2.6	3.1
Limestone	1.1 - 5.5	3.2	3.4 - 5.6	4.6	3.4

[a] Excluding flue gas reheat.

[b] Electrical energy consumption of the FGD installation as a percentage of gross.

FANS

Table 3-3 shows the trends in fan preference used on FGD systems. Although most of these fans are centrifugal, utilities are considering more innovative designs. Because early FGD systems were considered separate from the rest of the generating plant, separate booster fans provided draft for the scrubbing systems. Newer power plants have fans sized to provide draft for the entire boiler/scrubber installation as a unit. Where ESP's or baghouses provide particulate matter removal prior to the scrubbing system, forced-draft fans (with respect to the scrubber) are used extensively. These fans operate on dry flue gas. Most induced-draft (ID) fans operate on dry flue gas as well because they are often installed downstream from reheaters. Carbon steel is now and will continue to be the primary construction material for fans.

ABSORBERS

Table 3-4 is a breakdown of the number and capacity of units equipped with FGD systems according to generic absorber type and status. Combination absorbers include spray/packed and tray/packed absorbers as well as concentric venturi/spray tower absorbers. Impingement towers are fixed-baffle or fixed-vane

TABLE 3-3. NUMBER AND CAPACITY OF UNITS BY FAN SPECIFICATION AND INITIAL STARTUP YEAR

	Year of actual or projected FGD system initial startup					
	1971-1974		1975-1978		1979-1982	
Fan specification	No.	MW[a]	No.	MW[a]	No.	MW[a]
Design						
Centrifugal	8	2,198	31	12,529	34	12,880
Axial	0	0	1	185	4	1,313
NR[e]	3	145	1	200	37	14,315
Function						
Unit	3	191	21	9,623	13	5,417
Booster	5	1,199	11	2,849	22	7,482
NR[e]	3	945	1	442	40	15,609
Application[b]						
ID[c]	7	2,073	10	4,041	11	3,833
FD[d]	2	250	23	8,873	40	16,411
NR[e]	2	20	0	0	24	8,264
Service						
Wet	1	408	3	2,344	6	1,820
Dry	8	1,915	30	10,570	49	19,433
NR[e]	2	20	0	0	20	7,255
Materials						
Alloy	1	408	3	2,344	3	1,141
Carbon steel	8	1,915	28	9,850	47	18,443
Rubber-lined carbon steel	0	0	2	720	0	0
NR[e]	2	20	0	0	25	8,924

[a] Equivalent scrubbed capacity.
[b] With respect to the FGD system.
[c] Induced draft.
[d] Forced draft.
[e] Not reported.

absorbers, such as the disc contactor design. Fixed- or static-bed, mobile-bed, and rod-deck absorbers are considered packed towers. Systems in which flue gas is contacted with a slurry or solution such that the flue gas is adiabatically humidified and the slurry or solution is evaporated to apparent dryness are defined as spray dryers. Both horizontal and vertical spray absorber modules, which use radial, central, cocurrent, countercurrent, or crosscurrent spray arrangements, are considered spray towers. Impingement, sieve, and valve tray absorbers are considered tray towers. Fixed- and variable-throat venturi scrubbers as well as other absorber designs that operate on a venturi principle are grouped under venturi absorbers.

TABLE 3-4. NUMBER, CAPACITY, AND STATUS OF UNITS EQUIPPED WITH FGD SYSTEMS BY ABSORBER TYPE[a]

Absorber type	Operational		Under construction		Contract awarded	
	No.	MW[b]	No.	MW[b]	No.	MW[b]
Combination absorbers	10	3269	6	2871	3	1391
Impingement tower	1	265	0	0	2	842
Packed tower	19	6265	8	3211	2	750
Spray dryer	0	0	5	1660	6	1863
Spray tower	20	7181	16	7075	15	8008
Tray tower	15	4396	3	1802	1	65
Venturi absorber	8	3391	1	235	0	0

[a] These totals include SO_2 absorbers. Particulate matter scrubbers are excluded.

[b] Equivalent scrubbed capacity.

Table 3-4 indicates that spray towers have retained their popularity and that spray dryers will become more prominent in the 1980's. Except for venturis, which are on the decline, and these two prominent designs, the other absorbers show no marked change in commercial acceptability.

MIST ELIMINATORS

Utilities and system designers apparently prefer mist eliminators of the chevron design, particularly when they are preceded by a bulk separator. The primary material of construction is plastic, although some mist eliminators are made of alloys. None of those in the contract awarded status and only one unit now under construction will be constructed of materials other than plastic.

Most mist eliminators are horizontal, that is they are installed perpendicular to the vertically rising gas stream of conventional vertical absorbers. Vertical mist eliminators are used in horizontal absorber modules and some vertical absorbers that have a 90-degree turn of the duct (and thus a horizontal duct section before entry into the stack). The advantage of a vertical mist eliminator is that the liquid collected is removed perpendicular to the gas flow rather than opposite to it, thus improving the liquid removal efficiency. These patterns are somewhat evident in Table 3-5.

TABLE 3-5. NUMBER AND CAPACITY OF FGD-EQUIPPED UNITS BY MIST ELIMINATOR TYPE, CONFIGURATION, AND INITIAL STARTUP YEAR

	Year of actual or projected FGD system initial startup					
	1971 - 1974		1975 - 1978		1979 - 1982	
	No.	MW[a]	No.	MW[a]	No.	MW[a]
Type						
Chevron	10	2,202	34	12,106	38	14,929
Mesh-pad	1	110	1	360	0	0
Radial-vane	2	250	1	125	1	475
Configuration						
Horizontal	11	2,323	28	10,355	17	6,663
Vertical	0	0	6	1,418	5	1,793

[a] Equivalent scrubbed capacity.

REHEATERS

Four reheat strategies are currently in use or planned for domestic utility FGD systems: flue gas bypass, direct-combustion, hot-air-injection, and in-line reheat. In direct-combustion systems, fuel oil or gas is burned and hot combustion products are mixed with the wet scrubbed gas before it enters the stack. Hot-air-injection systems heat ambient air on the shell side of a steam tube heat exchanger and inject it into the flue gas stream. In-line reheaters heat the flue gas as it passes through the duct and contacts the reheater tubes. Both of the latter two systems use steam tubes with circulating steam or pressurized hot water for heat transfer. In some instances a unit will combine reheat systems. For example, where the percent of gas scrubbed can be made to vary with coal sulfur content, the flue gas is reheated by bypassing the particle-cleaned gas around the scrubbing system to the scrubber exit ductwork until the amount of allowable bypassed gas becomes inadequate for the required degree of reheat (when the percent sulfur is high), at which point a backup hot-air-injection reheater is activated.

Another variation of the basic reheater is the waste-heat recovery reheater. A waste-heat recovery reheater on a system currently under construction is an in-line reheater that includes two heat transfer areas. In the first transfer area, upstream of the scrubber, heat is absorbed from the flue gas; water circulating through heat exchanger tubes transfers the heat to a second transfer area downstream from the scrubber.

Table 3-6 is a breakdown of the reheat processes reported by number and capacity of units where these systems are installed or planned.

TABLE 3-6. NUMBER, CAPACITY, AND STATUS OF UNITS USING FLUE GAS REHEAT STRATEGIES

Reheat type	Operational		Under construction		Contract awarded	
	No.	MW[a]	No.	MW[a]	No.	MW[a]
Bypass	19	7,149	10	4,661	2	1,320
Bypass/hot air injection	1	447	1	447	0	0
Direct-combustion	10	2,589	1	240	0	0
Hot-air-injection	21	6,738	6	2,570	3	1,475
In-line	14	5,441	3	1,375	3	286
Waste-heat recovery	0	0	2	1,408	0	0

[a] Equivalent scrubbed capacity.

Five units (1687 MW) that are operational, one unit (110 MW) that is under construction, and five units (1416 MW) for which contracts have been awarded do not include reheaters.

STACK FLUES

Table 3-7 is a breakdown of units according to materials of construction of the stacks, status, and whether or not they have reheat. The flues of most stacks are and continue to be made of

TABLE 3-7. NUMBER, CAPACITY, AND STATUS OF UNITS EQUIPPED WITH FGD SYSTEMS ACCORDING TO FLUE/LINER TYPE AND REHEAT APPLICATION

Flue/liner	Operational				Under construction				Contract awarded			
	With reheat		Without reheat		With reheat		Without reheat		With reheat		Without reheat	
	No.	MW[a]	No.	MW[a]	No.	MW[a]	No.	MW[a]	No.	MW[a]	No.	MW[a]
Alloy	0	0	1	917	0	0	0	0	0	0	0	0
ARBM[b]	19	6103	6	2015	11	5472	4	1455	4	1426	3	1687
Carbon steel	5	2976	0	0	0	0	0	0	0	0	0	0
C.S.[c]/inorganic lining	2	1834	1	98	0	0	0	0	0	0	0	0
C.S.[c]/organic lining	7	2369	2	514	0	0	0	0	0	0	0	0
Fiberglass	2	455	0	0	3	1220	0	0	2	1000	0	0
HCBC[d]	10	2370	0	0	1	242	0	0	0	0	0	0

[a] Equivalent scrubbed capacity.
[b] Acid resistant brick and mortar.
[c] Carbon steel.
[d] Hydraulic-cement-bonded concrete.

acid-resistant brick and mortar (ARBM). Information regarding materials of construction in the units under construction or on which contracts have been awarded is lacking partially because utilities often do not finalize stack design until late in the construction stage.

SLUDGE DISPOSAL

Table 3-8 is a breakdown of units equipped with FGD according to sludge treatment, transportation, disposal method, site, and operational status. As in the case of stacks, information on units under construction and on which contracts are awarded is incomplete because final disposal strategies are often not finalized until plant construction is nearly complete. Also, when a separate contract is arranged for sludge disposal, it is often not awarded until after initial plant construction.

Most disposal sites are and will continue to be on the plant site. One trend is to increase sludge solids content by fly ash addition and/or using vacuum filters so the material can be landfilled. Another trend is to provide some sort of sludge treatment before final disposal; primary methods are fly ash stabilization, forced oxidation, and proprietary fixation. As more systems produce sludge with higher solids content, waste transport by truck and/or conveyor belt will become more prominent.

TABLE 3-8. NUMBER, CAPACITY, AND STATUS OF UNITS EQUIPPED WITH FGD SYSTEMS ACCORDING TO SLUDGE DISPOSAL SPECIFICATIONS AND STATUS

Disposal specification	Operational		Under construction		Contract awarded	
	No.	MW[a]	No.	MW[a]	No.	MW[a]
Sludge treatment type						
Bottom ash addition[b]	1	490	0	0	0	0
Fly ash/lime stabilization	5	956	0	0	1	65
Fly ash addition[c]	9	3,494	0	0	0	0
Fly ash mixing[d]	3	1,785	3	1,219	2	1,000
Forced oxidation	4	2,025	6	3,430	1	166
Proprietary fixation	11	5,615	6	2,686	2	1,370
Sludge transportation						
Conveyor	4	1,070	2	1,140	0	0
Pipeline	29	10,666	1	280	0	0
Rail	3	1,785	1	500	0	0
Truck	12	3,526	6	2,733	5	2,146
Sludge disposal method						
Landfill	21	9,011	16	7,858	10	3,824
Lined pond	30	9,408	9	3,943	1	50
Mine fill	2	632	3	1,421	0	0
Unlined pond	14	3,971	0	0	0	0
Disposal site						
Onsite	55	15,915	17	8,002	2	1,067
Offsite	14	4,899	4	1,397	2	120

[a] Equivalent scrubbed capacity.
[b] FGD wastes and bottom ash are disposed of together.
[c] FGD wastes and fly ash are disposed of together.
[d] FGD wastes and fly ash are mixed before final disposal.

SECTION 4

PERFORMANCE TRENDS

OPERATING EXPERIENCE

In the past 5 years, FGD has become the most commercially developed means of control of SO_2 emissions from coal-fired boilers, and operating experience has increased significantly. At the end of 1975, 20 units were either on line (or had been), and approximately 198,000 hours of on-line experience had been accumulated. By August 1980, 85 FGD systems had been operated on utility boilers, and more than 460,000 hours of operation had been logged. This represents a 425% increase in the number of FGD systems operated and a 230% increase in total hours logged.

The operational hours above reflect the number of hours reported by the utilities. Because hours of operation often are not available for such periods as initial system startup or performance testing, the actual number of operational hours is greater than reported, as is the corresponding percentage increase.

DEPENDABILITY

For characterization of system performance, four dependability parameters have been developed: availability, operability, reliability, and utilization. Table 4-1 defines these parameters.

The FGD survey program includes monitoring the performance of the operating FGD systems and logging monthly operating parameters (e.g., boiler and FGD system operating hours, forced outage times, scheduled outage times). If the data permit, monthly dependability parameters are calculated for both the entire FGD system and its respective modules (where applicable). When modular operating parameters are known, total FGD system dependability parameters are derived by averaging all the modular figures, except in those cases where the FGD system design includes spare capacity. In these instances, a spare capacity factor is included in the calculation of the total system parameter, which ensures that the overall FGD system dependability is not penalized as a result of equipment redundancy.

TABLE 4-1. PARAMETERS OF DEPENDABILITY

Availability index	Hours the FGD system is available for operation (whether operated or not) divided by the hours in the period.
Operability index	Hours the FGD system was operated divided by the boiler operating hours in the period.
Reliability index	Hours the FGD system was operated divided by the hours it was called upon to operate.
Utilization index	Hours the FGD system operated divided by the total hours in the period.

Figures 4-1 and 4-2 reflect the availability history of four FGD installations on boilers firing low- or medium-sulfur ($\leq$2.5%) coal, and Figures 4-3 and 4-4 show the availability of four FGD installations on boilers firing high-sulfur (>2.5%) coal. These units represent systems for which sufficient data are available for analysis. In each case, the data points represent 12-month rolling averages of the monthly total system availabilities. The rolling averages are calculated by averaging the availability data for the first 12 months of operation, dropping the first data point, and adding the 13th for a second average, and so on.

Figures 4-5, 4-6, and 4-7 are composites of the availabilities of individual FGD systems. They show average annual availabilities (through June 1980) for operating units firing low-sulfur (<1%) coal, medium-sulfur (1-3%) coal, and high-sulfur (>3%) coal, respectively.* Some newly operational systems were not included (even though data were available) because they had been in operation for less than 1 year and yearly availability averages were not available.

Figure 4-8 provides statistical analyses of the data contained in the three composite graphs for 1978 and 1980. In each case, the availability points for these two years were plotted, and the median of each array was determined. Note that the median FGD system availability for those systems applied to units firing high-sulfur (>3%) coal has shown a better than 50% increase in the 2-year period, and is approaching that of the low- to medium-sulfur coal units. This indicates a rising trend in the overall dependability of FGD systems for high-sulfur coal application. The median availability for units firing medium-sulfur (1-3%) coal increased 16.5% and, for units firing low-sulfur coal, 1.5%. The lower percent change for these two categories is attributable to their higher median availability in 1978 and attests to the stable and reliable operating histories experienced by FGD systems on these low- and medium-sulfur coal units.

SO_2 REMOVAL EFFICIENCY

Table 4-2 presents SO_2 removal efficiency performance test results and total system design removal efficiency values for some of the operational FGD systems. Table 4-3 presents continuous monitoring data for some of these systems. All but two of the systems represented in these tables are commercial lime/

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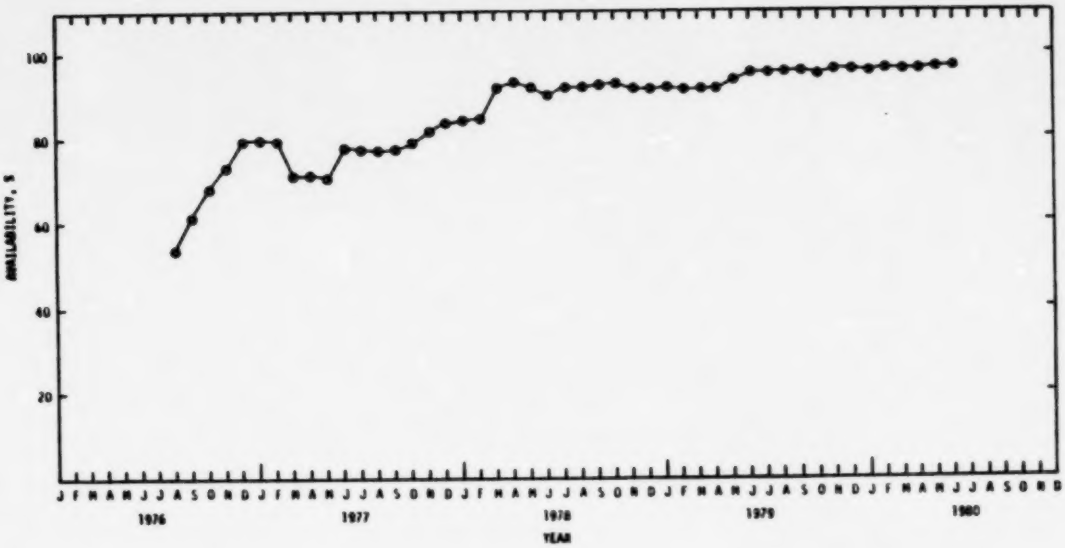
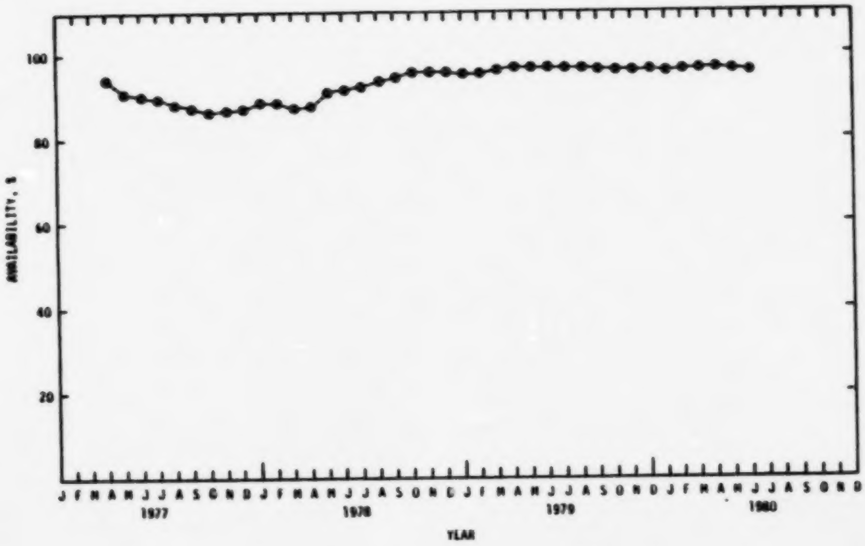
* These categories were used to provide a more even graphical distribution; however, they differ slightly from those used in previous sections.

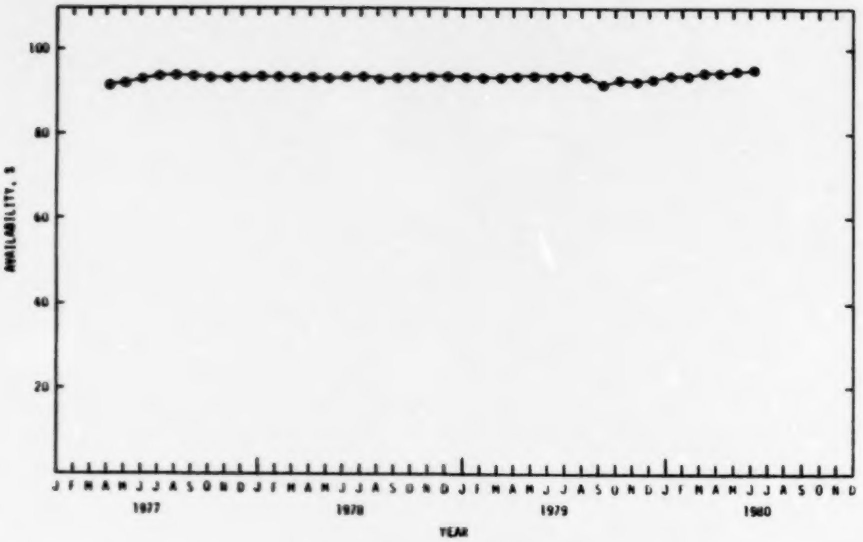


Colstrip 1

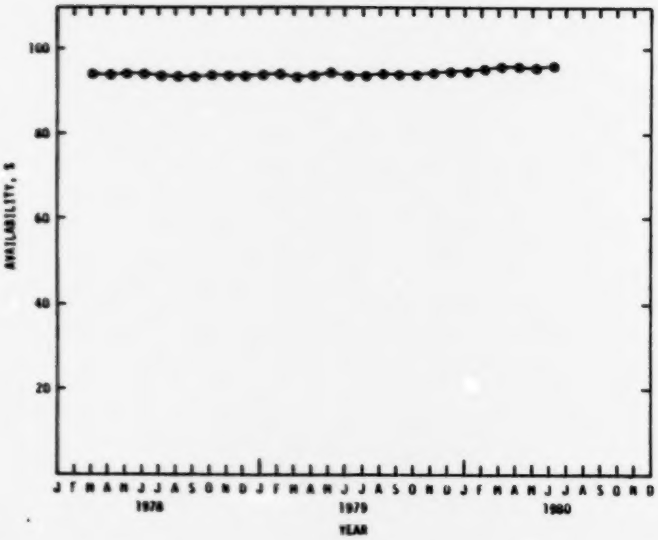


Colstrip 2

Figure 4-1. Availability histories for FGD installations at the Colstrip Station of Montana Power (≤2.5% S coal).

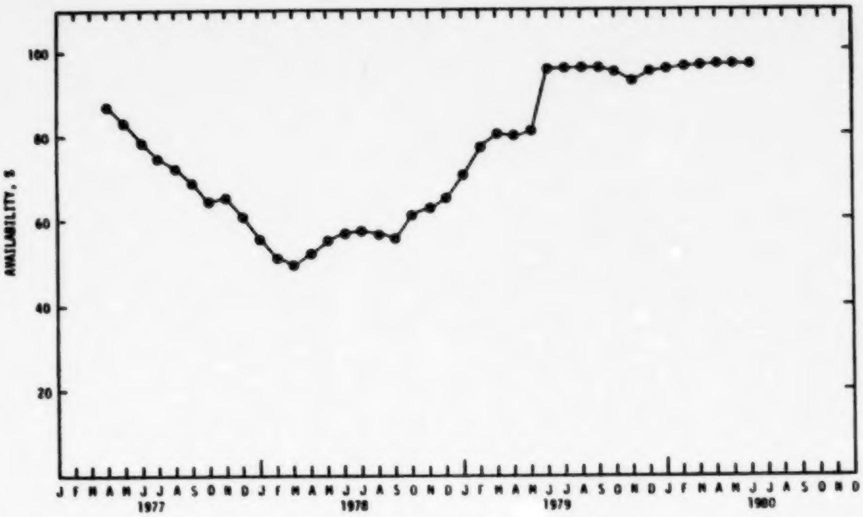


Sherburne 1

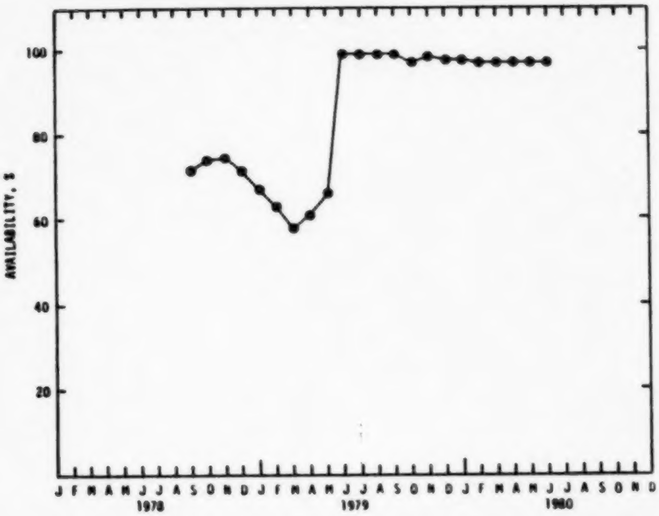


Sherburne 2

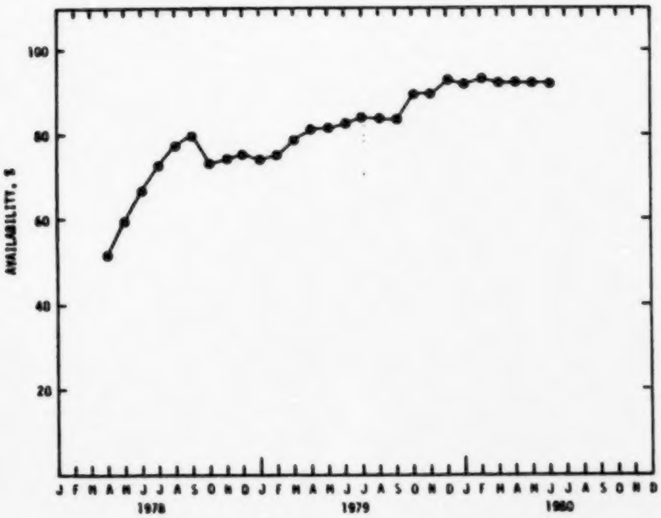
Figure 4-2. Availability histories for FGD installations at the Sherburne Station of Northern States Power ($\leq$2.5% S coal).



Bruce Mansfield 1



Bruce Mansfield 2



Widows Creek 8

Figure 4-3. Availability histories for FGD installations at the Bruce Mansfield Station of Pennsylvania Power and Widows Creek Station of Tennessee Valley Authority (>2.5% S coal).

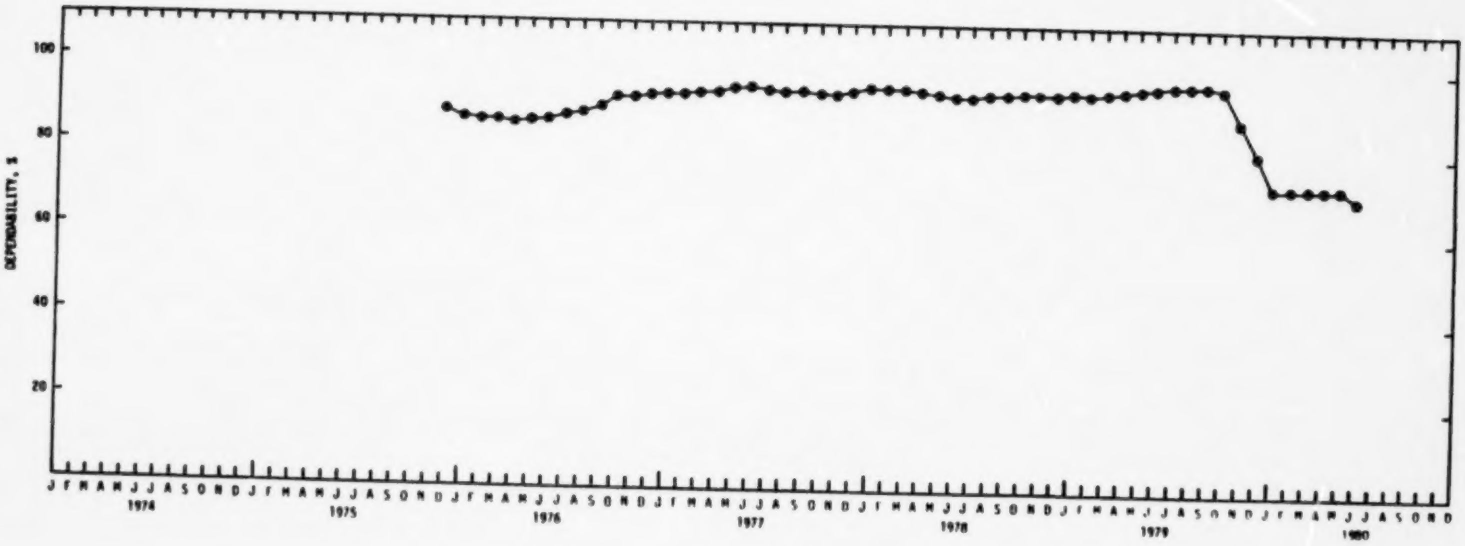


Figure 4-4. Availability history for the FGD Installation at the LaCygne Station of Kansas City Power and Light (>2.5% S coal).

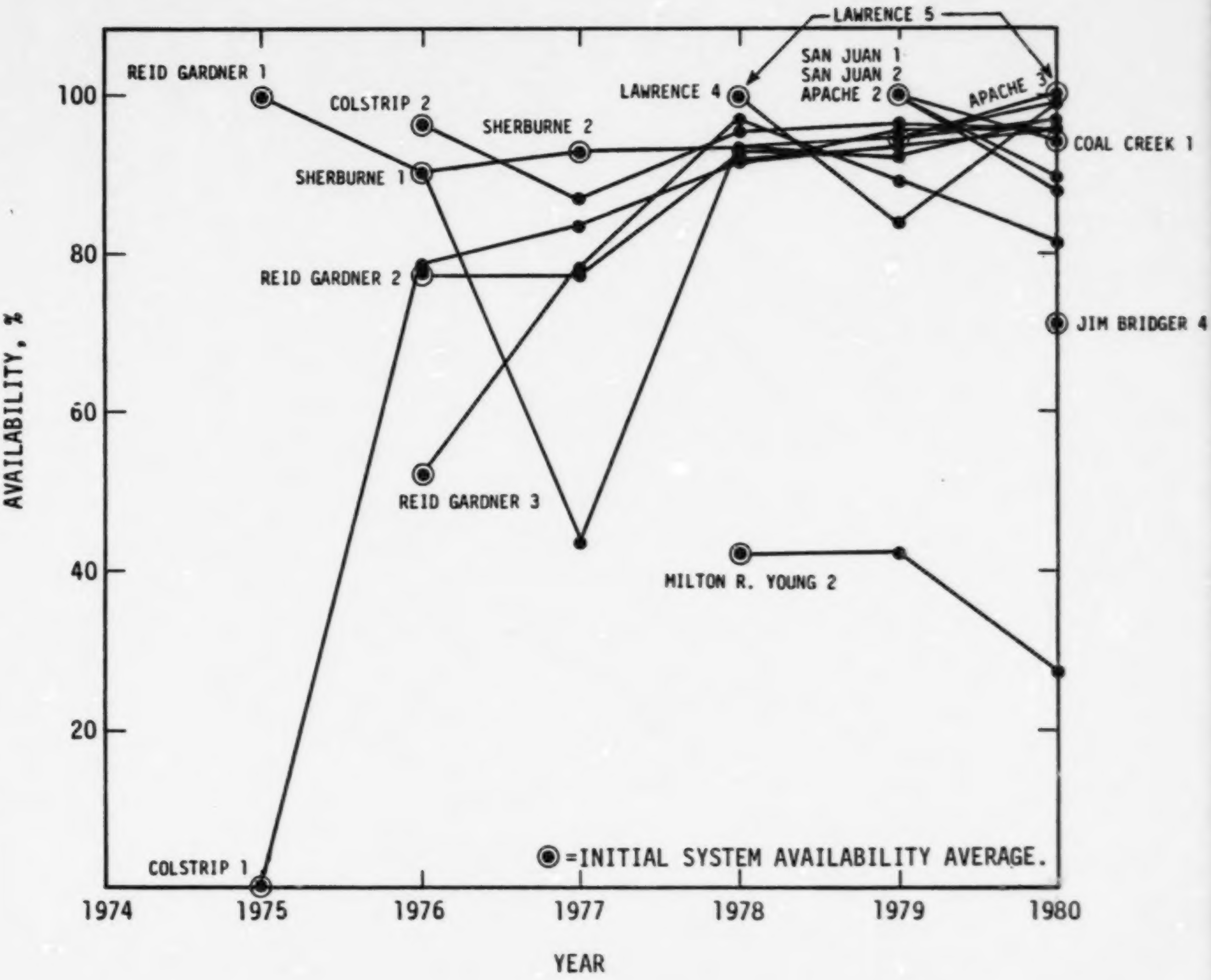


Figure 4-5. Annual average availability histories for low sulfur (<1%) coal FGD installations.

88%

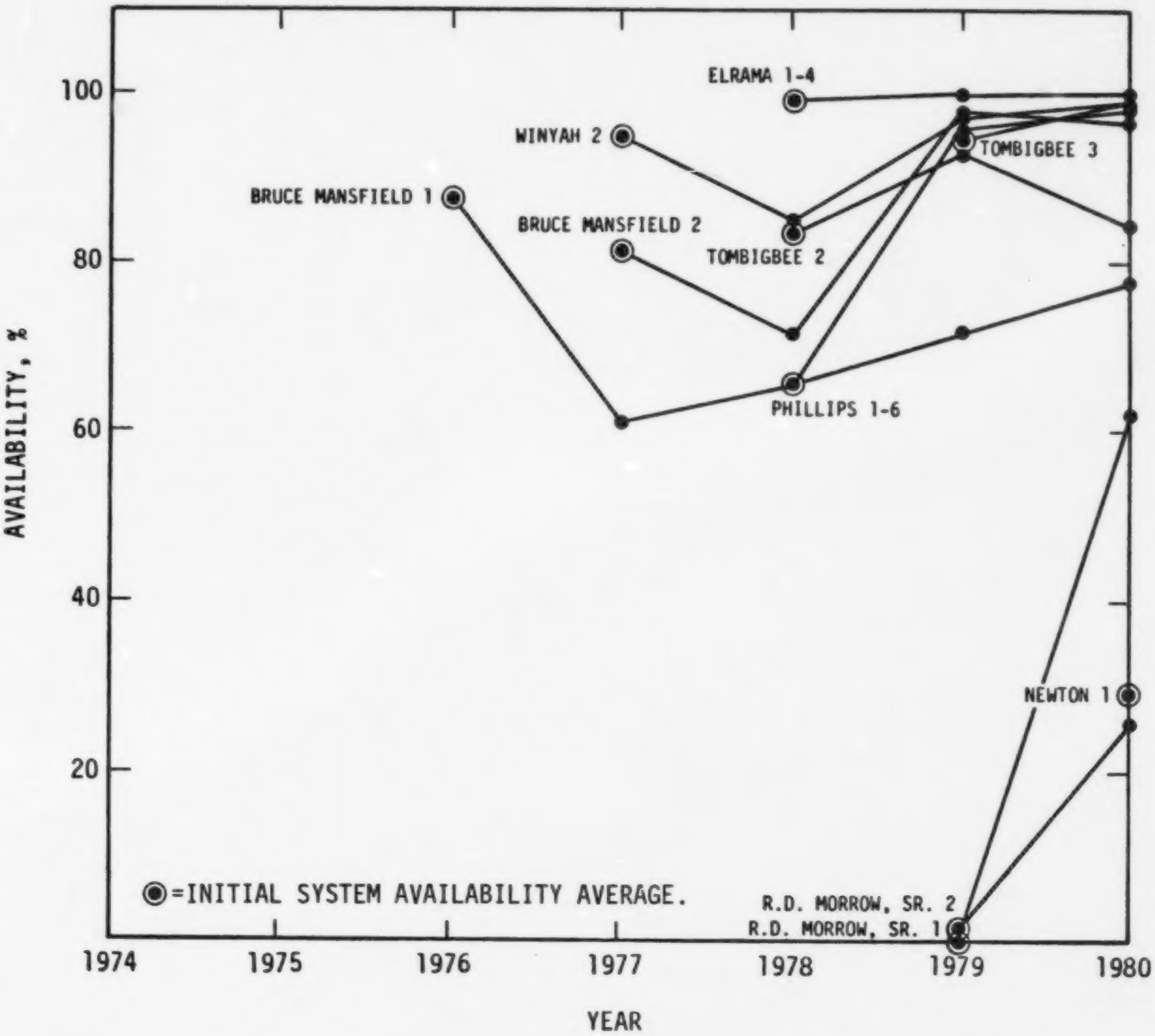


Figure 4-6. Annual average availability histories for medium sulfur (1-3%) coal FGD installations.

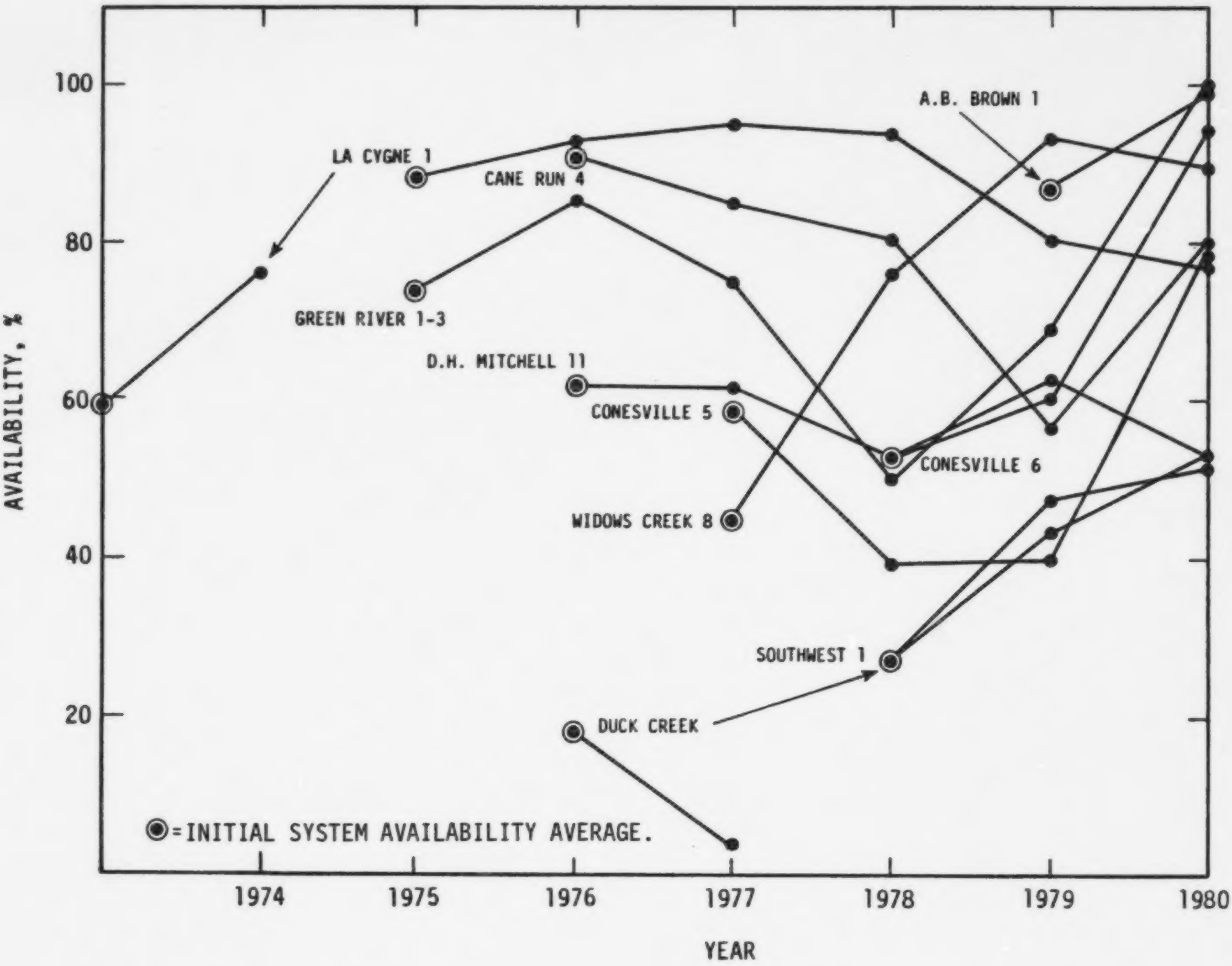
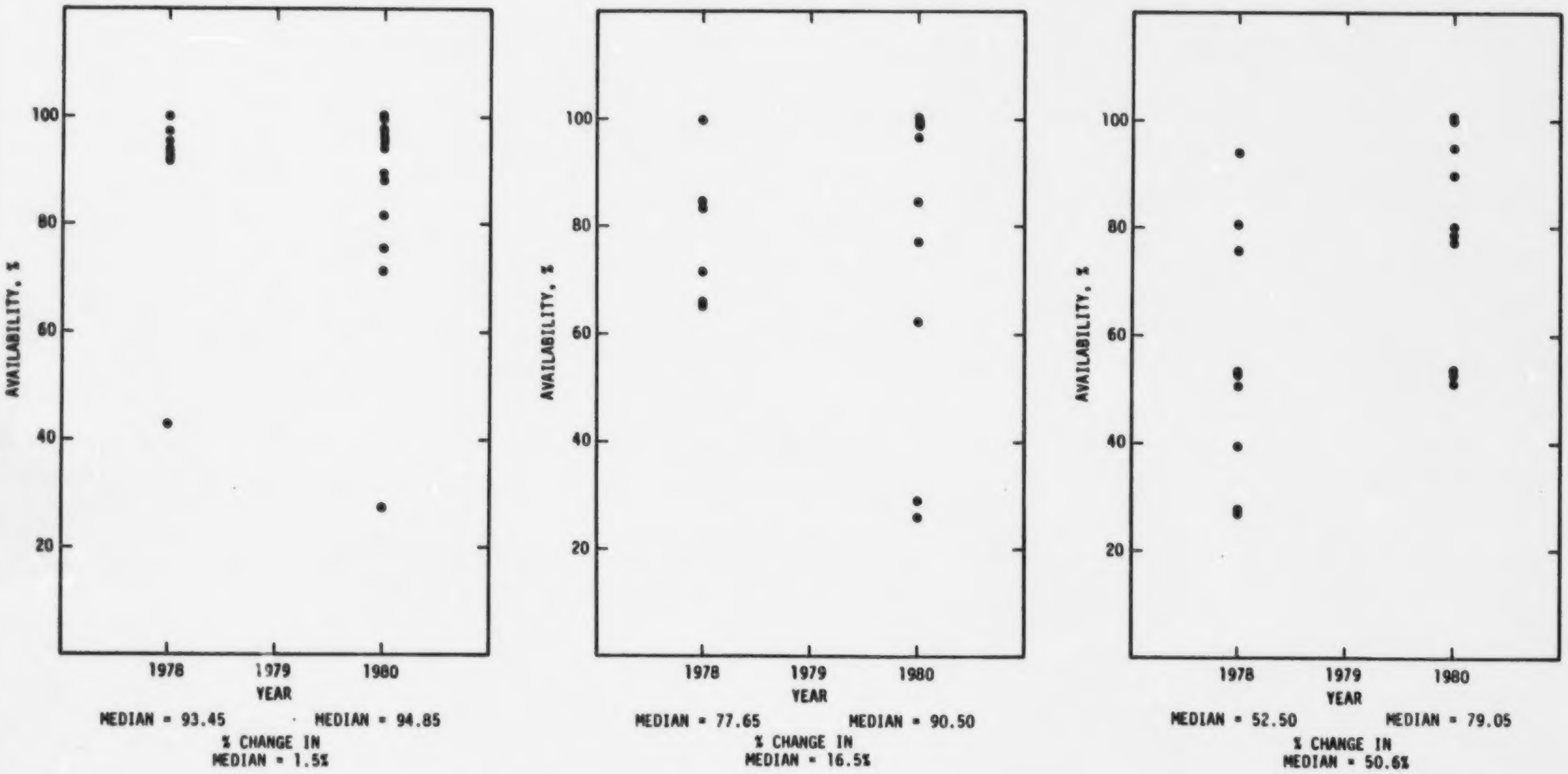


Figure 4-7. Annual average availability histories for high sulfur (>3%) coal FGD installations.



Low sulfur coal installations. Medium sulfur coal installations. High sulfur coal installations.

Figure 4-8. Statistical analyses of the annual availability data for the years 1978 and 1980.

TABLE 4-2. SO_2 REMOVAL EFFICIENCIES; PERFORMANCE TEST DATA

Utility name/ unit name	Unit rating, MW (gross)	Process type	Fuel sulfur content, %	Design removal efficiency, %	Date	Performance test results, %	Remarks
Arizona Public Service Cholla 1	119	Limestone	0.5	92[a]	10/73	92	Test results are based on testing of Module A only
					10/73	58.5	Test results are based on the average of tests from October 2, to October 21, 1973
Duquesne Light Phillips 1-6	408	Lime	1.5	83[b]	1975	86-93	Tests results are from two-stage scrubbing train
Kansas City Power & Light LaCygne 1	874	Limestone	5.4	80	3/75	77	Test results were taken from a 4-hour full load test
					5/75	80	Results are based on an 8-hour maximum continuous load test
					8/77	77	Summary of a 4-hour full load test
Kansas Power & Light Lawrence 4	125	Limestone	0.6	73	10/77	96-98	Summary of overall results from acceptance tests
Kentucky Utilities Green River 1-3	64	Lime	4.0	80	10/78	83	Results are the average of six test runs
Louisville Gas & Electric Can Run 4	188	Lime	3.8	85	3/77	95	Results of a 7- to 10-day test period
					8/77	86-89	Performance test results
Can Run 5	200	Limestone	3.8	85	7/79	88	The result is an average of three emission tests
Cane Run 6	299	Dual alkali	4.8	95	7/80	94	The result is from compliance test performed over an 11-day period
Montana Power Colstrip 1	360	Lime/alkaline flyash	0.8	60	2/76	75	Tests were EPA Method 6 procedures
					1/77	81	
					5/77	88	
					6/77	81	
Colstrip 2	360	Lime/alkaline flyash	0.8	60	10/76	68	Tests were EPA Method 6 procedures
					11/76	83	
					12/76	83	
					3/77	86	
					6/77	83	

(continued)

TABLE 4-2 (continued)

Utility name/ unit name	Unit rating, MW (gross)	Process type	Fuel sulfur content, %	Design removal efficiency, %	Date	Performance test results, %	Remarks
Northern Indiana Public Service D.H. Mitchell 11	115	Wellman Lord	3.5	90	9/77	91	Tests commenced on Aug. 29, 1977, and were completed on Sept. 14, 1977; test period included 12 days at 92 MW flue gas equivalent and 3-1/2 days at 110 MW flue gas equivalent
South Mississippi Elect. Power R.D. Morrow, SR. 1	200	Limestone	1.3	85[c]	3/80	92	Results of five EPA Method 6 tests across the absorber
					4/80	90	Results of seven EPA Method 6 tests across the absorber
Springfield City Utilities Southwest 1	194	Limestone	3.5	80	9/77	92	Average result of compliance test runs
Texas Utilities Martin Lake 1	793	Limestone	0.9	95[d]	6/77	98-99	Preliminary acceptance test results at 750 MW
					8/78	98-99	Acceptance test results

[a] Module A removal efficiency; overall unit design removal efficiency is 59%.
[b] Design removal efficiency of the two-stage scrubbing trains.
[c] Absorber design removal efficiency; overall removal efficiency is 53%.
[d] Absorber design removal efficiency; overall removal efficiency is 71%.

TABLE 4-3. SO_2 REMOVAL EFFICIENCIES: CONTINUOUS MONITORING DATA

Utility name/ unit name	Unit rating, MW (gross)	Process type	Fuel sulfur content, %	Design removal efficiency, %	Date	Actual removal efficiency, %
Colorado Ute Craig 2	455	Limestone	0.5	85	5/80	65
					6/80	66
					7/80	66
					8/80	66
Kansas City Power & Light LaCygne 1	874	Limestone	5.4	80	9/77	81
Kansas Power & Light Lawrence 4	125	Limestone	0.6	73	10/77	97
					2/79	94
Louisville G&E Cane Run 4	188	Lime	3.8	85	7/77	81
					8/77	84
					10/77	84
					11/77	84
					7/80	87
Cane Run 5	200	Lime	3.8	85	7/80	85
Cane Run 6	288	Dual alkali	4.8	95	6/80	95
Mill Creek 3	442	Lime	3.8	85	6/80	85
Montana Power Colstrip 1	360	Lime/alkaline flyash	0.8	60	4/76	86
					7/76	90
					9/76	89
					12/76	81
Northern Indiana Public Service D.H. Mitchell 11	115	Wellman Lord	3.5	90	8/77	90
					10/77	90
					11/77	91
Northern States Power Sherburne 2	740	Limestone/ alkaline flyash	0.8	50	4/77	58
Pennsylvania Power Bruce Mansfield 1	917	Lime	3.0	92	10/77	81
Philadelphia Electric Eddystone 1A	120	Magnesium oxide	2.6	90	9/77	97
					11/77	85
South Carolina Public Service Winyah 2	280	Limestone	1.7	69	6/79	80
					7/79	84
					8/79	80
South Mississippi R.D. Morrow, SR. 1	200	Limestone	1.3	85[a]	4/80	80
					5/80	80
					6/80	90
					7/80	90
					8/80	80
R.D. Morrow, SR. 2	200	Limestone	1.3	85[a]	9/79	95
					5/80	85
					6/80	90
					7/80	85
					8/80	80

(continued)

TABLE 4-3 (continued)

Utility name/ unit name	Unit rating, MW (gross)	Process type	Fuel sulfur content, %	Design removal efficiency, %	Date	Actual removal efficiency, %
Tennessee Valley Authority Widows Creek 8	516	Limestone	3.7	89	11/77	91
					12/77	94
					1/78	89
					2/78	85
					3/78	92
					4/78	90
					5/78	89
					6/78	92
					7/78	88
					8/78	89
					9/78	91
					5/79	80
					6/79	84
					7/79	86
					8/79	88
					9/79	83
					10/79	87
					11/79	88
					12/79	86
					1/80	84
					2/80	84
					3/80	83
					4/80	86
					5/80	83
					6/80	82
					7/80	87

[a] Absorber design removal efficiency; overall removal efficiency is 53%.

limestone installations. The two exceptions are demonstration systems utilizing dual alkali and magnesium oxide processes. The available data, although not extensive, indicate that actual removal efficiencies of these systems generally meet or exceed design values at both low-sulfur and high-sulfur coal installations. This would seem to indicate that meeting or exceeding design SO_2 removal efficiency has not been a significant problem for FGD systems on units firing high-sulfur coal. For example, the FGD installation at the La Cygne power station (the FGD-equipped unit currently firing the highest-sulfur coal) successfully passed performance testing early in 1975. Results from 10 days of continuous monitoring in late September 1977 indicated that the system was continuing to exceed its design removal efficiency of 80%.

PERFORMANCE CONSIDERATIONS

Because of the widely varying conditions at stations where FGD systems are applied (e.g., differences in plant size, coal sulfur content, and required removal efficiencies), it is difficult to pinpoint specific variables affecting overall FGD system performance. Certain general considerations can be identified, however, and are discussed below.

SO_2 Inlet Levels and Removal Requirements

In general, FGD systems operating on units with low to medium SO_2 inlet levels have demonstrated a higher level of overall dependability than those operating on units with higher inlet levels. This is illustrated in the statistical analyses of the overall FGD system availability (Figure 4-8) for low-sulfur coal units. Obviously, the lower SO_2 removal requirement contributes to this difference.

Unit Load Profile and Coal Characteristics

Higher dependabilities have resulted from a reduction in the number of chemical and mechanical problems on FGD systems applied to new, base-loaded boilers designed to fire coal from one or several specific sources. The flue gas generated by such units generally has more relatively constant and stable characteristics, and overall system dependability apparently improves because the system does not have to respond to dramatic variations in flue gas flow rates and composition. In FGD systems retrofitted to cycling, and peak-load units, these systems often must respond to conditions that reach or exceed their process control capabilities, and problems result from the variations that occur in reagent feed rate and loss of chemical control.

System Redundancy and Bypass Capability

FGD systems are now considered an integral part of the power generating plant, and more stringent regulations prevent many utilities from bypassing the FGD system. Thus, the current design trend is toward incorporation of spare absorber modules and ancillary equipment. Systems so designed have greater dependability because the failure of a single component does not necessarily force the entire system off line. Spare capacity also promotes a more flexible operating and maintenance strategy by allowing some routine maintenance to be performed without removing the system from service. The result is an overall reduction in FGD system downtime.

Utility Experience

As utilities continue to gain more experience with FGD system operation, overall system dependabilities are expected rise. In the early stages of FGD operation, utility staffs had little experience with the chemical processes involved in FGD operation, and the chemical and mechanical problems that are inevitable with complex processes such as these were difficult to rectify. The steadily increasing commercial operating hours will allow system operators and maintenance personnel to gain the experience necessary for more efficient and expeditious analysis of system problems and implementation of solutions. In addition, utilities are employing more chemical engineers and other personnel familiar with gas/liquid systems to deal with these problems.

Operating and Maintenance Philosophy

A general trend in plant philosophy regarding operation and maintenance (O&M) is the dedication of specific crews to handle this responsibility, rather than considering it a secondary function of the power plant O&M personnel. This change will permit faster and more precise changing of system parameters to meet varying load conditions, and overall system reliability should improve as problems are attended to expeditiously.

System Design Generation

Building on experience gained in the operation of first-generation systems, system suppliers and designers are now providing better process design configurations and materials of construction. Indicative of this trend are the broader guarantees system suppliers are now offering with respect to SO_2 removal efficiency, mist carryover, waste stream quality/quantity, power consumption, reagent consumption, and availability. Many of the newer systems should exhibit fewer of the traditional operating problems, especially during the critical startup and debugging phases of operation.

SECTION 5

CAPITAL AND ANNUAL COSTS

INTRODUCTION

Another important function of the utility survey program is the acquisition and analysis of cost data. In this program, emphasis is on costs associated with operational systems because of the availability of meaningful and complete data. These data are adjusted only to ensure their completeness and accuracy and to facilitate comparison. The approach and methodology used in analyzing these costs and the results of these analyses are briefly described in the following subsections.

APPROACH

Capital and annual cost data on operational FGD systems have been obtained continuously since March 1978. Costs for each system are obtained directly from the utilities and from published sources, and then itemized by individual FGD cost element. The itemized costs are then adjusted to a common basis to enhance comparability. This adjustment includes factors for estimating costs not given by the utilities and escalating all costs to common dollars (mid-1980). All adjusted cost data and computations are reviewed and verified with the appropriate utility.

It is important to note that the costs analyzed here are real costs, not cost model projections. When a particular itemized cost is not reported by the utility, an estimated cost based on known system design and operating factors is included. The use of estimates is not arbitrary; they are used only when cost items are unavailable or are judged to be unrepresentative.

ADJUSTMENT PROCEDURE

Capital Costs

- All costs associated with control of particulate matter emissions are deducted.

- Capital costs for modifications necessitated by installation of an FGD system are added if they were not included in the reported costs.

- Sludge disposal costs are adjusted to reflect a 20-year life span for retrofit systems and a 30-year life span for new systems.

- Any unreported direct and indirect costs incurred are estimated and included.

- All capital costs are escalated to mid-1980 dollars.

- All $/kW values reflect the gross generating capacity of the unit.

Annual Costs

- All costs are adjusted to a common 65% capacity factor.

- Direct costs that were not reported are estimated and added.

- Overhead and fixed costs that were not reported are estimated and added.

- All annual costs are escalated to mid-1980 dollars.

- All mills/kWh values are based on a 65% capacity factor and the net generating capacity of the unit.

RESULTS

Table 5-1 summarizes both reported and adjusted costs for all 45 operational FGD systems on which cost data were obtained. This table also summarizes the results by application (new/retrofit) and by sulfur content of the coal (high sulfur/low sulfur). Table 5-2 lists the results by process type. A plant-by-plant listing of the reported and adjusted costs for the operational FGD systems addressed in this study is provided in Appendix B.

COST MODEL COMPARISON

During the past few years, various organizations have conducted major cost studies of the capital and annual costs associated with different FGD processes. Reasons for these cost studies range from comparing the economics associated with

TABLE 5-1. CATEGORICAL RESULTS OF THE REPORTED AND ADJUSTED CAPITAL AND ANNUAL COSTS FOR OPERATIONAL FGD SYSTEMS

	Reported				Adjusted			
	Capital		Annual		Capital		Annual	
Results	Range, ($/kW)	Average, ($/kW)	Range, (mills/kWh)	Average, (mills/kWh)	Range, ($/kW)	Average, ($/kW)	Range, (mills/kWh)	Average, (mills/kWh)
All	23.7-174.8	78.9	0.29-13.02	2.97	35.1-258.9	116.2	1.80-18.64	7.64
New	23.7-174.8	78.4	0.29- 5.81	2.19	35.1-242.1	107.4	1.80-13.44	6.49
Retrofitted	29.3-157.4	79.6	0.46-13.02	4.54	57.5-258.9	131.4	4.36-18.64	9.38
High sulfur	29.3-157.4	75.1	0.92-13.02	3.71	57.5-233.6	106.3	3.70-18.37	7.48
Low sulfur	23.7-174.8	82.3	0.29-11.32	2.09	35.1-258.9	122.6	1.80-18.64	7.40

TABLE 5-2. ADJUSTED CAPITAL AND ANNUAL COSTS FOR OPERATIONAL FGD SYSTEMS BY PROCESS TYPE

Process	Reported				Adjusted			
	Capital		Annual		Capital		Annual	
	Range ($/kW)	Average ($/kW)	Range (mills/kWh)	Average (mills/kWh)	Range ($/kW)	Average ($/kW)	Range (mills/kWh)	Average (mills/kWh)
Limestone	23.7-168.0	68.8	0.29- 7.80	2.47	35.1-148.7	99.6	1.80- 8.56	6.02
Lime	29.3-122.8	71.0	0.92-11.32	3.69	57.5-192.7	104.5	3.70-10.82	6.91
Dual alkali	47.2-174.8	97.8		1.30	80.6-242.1	134.6	5.10-13.44	8.11
Lime/ alkaline flyash	92.5-101.4	98.4	1.25- 2.97	2.40	131.0-133.8	132.9	5.99- 7.79	7.19
Sodium carbonate	42.9-113.6	72.4	0.23- 0.46	0.38	79.9-138.5	101.7	5.29- 6.78	6.02
Wellman-Lord	132.8-157.4	142.4		13.02	233.6-258.9	249.1	17.86-18.37	18.10
Limestone/ alkaline flyash		49.3		0.75		94.5		4.63

commercial and emerging FGD processes to determining the cost impact of increasingly stringent SO_2 standards. Table 5-3 presents the results of several representative cost studies recently completed and the assumptions on which they are based.

In this table, the capital and annual cost estimates and their underlying assumptions are summarized for a number of "base cases." In this context, "base case" refers to a conventional wet limestone slurry FGD process such as that typically installed on a new 500-MW (net) boiler firing high sulfur eastern coal. This table shows that capital and annual costs vary widely, with the capital values ranging from $94.5 to $194.4/kW and the annual values ranging from 4.03 to 16.91 mills/kWh. These wide variations in estimated costs for essentially the same case result from differences in the intent of the studies and in the assumptions on which each is based. With respect to the latter, variations can be noted for virtually every key assumption.

By use of the reported and adjusted capital and annual costs for the operational FGD systems presented in Appendix B, it was possible to compare the estimated costs in these cost studies with actual costs. For this comparison, only limestone systems have been analyzed, as this was the "base case" of all the aforementioned cost studies.

Table 5-4 presents the adjusted capital and annual costs of the limestone systems currently in service on coal-fired utility boilers. Generally, these costs represent the technology of first-generation limestone systems that have been operational for several years. Many have bypass capabilities. Most of these systems scrub less than 100% of the flue gas and therefore do not require a separate reheat system. A significant number of units have total removal efficiencies of less than 70%. Few systems have spare components and few have oversized components to provide spare capacity. Sludge is typically disposed of in ponds without fixation or treatment.

A comparison shows that capital and annual costs of actual systems approach the costs developed by the Tennessee Valley Authority (TVA) and Beychok cost studies. The actual average capital cost for limestone FGD systems is $99.6/kW; average annual cost is 6.03 mills/kWh. The TVA cost study arrived at a capital cost of $97.5/kW and an annual cost of 4.03 mills/kWh; the Beychok cost study, a capital cost of $94.5/kW and an annual cost of 6.61 mills/kWh. The criteria used in developing the costs in these two studies are also based on early FGD technology.

Assumptions used in the other cost studies reflect future, more advanced FGD system designs. They also reflect inclusion

TABLE 5-3. BASE CASE CAPITAL AND ANNUAL COST ESTIMATES

Category	FGD economic studies					
	Bechtel[3]	PEDCo Environmental[4]	Stearns-Roger[5]	Beychok/ Stone & Webster[6]	SRI/Radian[7]	TVA[8]
General criteria:						
Sponsoring organization	EPRI	EPA	EPRI	EPRI	EPRI	EPA
Year prepared	1979	1977	1979	1977 to 1978	1979 to 1980	1979
Plant size	2-unit	1-unit	1-unit	1-unit	1-unit	1-unit
Plant location	North Central	Midwest	North Central	Midwest	NA	Midwest
Plant capacity, MW (net)	1000	500	500	500	499	500
Plant capacity factor, %	70	65	70	70	70	80
Plant application	New	New	New	New	New	New
Plant heat rate, Btu/kWh (net)	9986	9000	9724	9000	NA	9000
Fuel (source)	Coal (Illinois)	Coal (Eastern)	Coal (Illinois)	Coal (Eastern)	Coal (Eastern)	Coal (Eastern)
Fuel characteristics, HV/S%/A%	10,100/4.0/16.0	12,000/3.5/14.0	10,100/4.0/16.0	12,000/3.5/NA	10,100/4.0/16.0	10,500/3.5/16
Emission standard	Revised NSPS[a]	Revised NSPS[b]	Revised NSPS[c]	Revised NSPS[d]	Revised NSPS[e]	NSPS[f]
SO_2 emissions, lb/10^6 Btu	1.2	0.6	0.8	0.5	0.5	1.2
SO_2 removal efficiency, %	87	90	90	90	93	80
Process design criteria:						
Process	Limestone	Limestone	Limestone	Limestone	Limestone	Limestone
Number of modules	8	5	4	NA	5	4
Number of spares	2	1	1	NA	1	0
Gas bypass capability	Complete bypass	Complete bypass	Complete bypass	NA	Complete bypass	No bypass
Reheat ΔT, °F	56	50	50	Yes	50	50
Water loop	Closed	Closed	Closed	Closed	Closed	Closed
Solids dewatering, % solids	50	50	45	Yes	60	None
Sludge treatment	Fly ash/lime	Fly ash/lime	Fly ash/lime	Fly ash/lime	Fly ash/lime	None
Sludge disposal	Truck/landfill	Pumping/pond	Truck/landfill	NA	Truck/landfill	Pumping/pond
Economic criteria:						
Capital cost basis	Total	Total	Total	Total	Total	Total
Annual cost basis	Total first year revenues	Total first year revenues	Total first year revenues	Total first year revenues	Total first year revenues	Total first year revenues
Battery limits	Gas inlet to sludge disposal	Gas inlet to sludge disposal	Gas inlet to sludge disposal	Gas inlet to sludge disposal	Gas inlet to sludge disposal	Gas inlet to sludge disposal
Price level	July 1978	July 1980	July 1978	First quarter 1977	January 1979	Mid-1979 (capital) Mid-1980 (annual)
Cost estimates:						
Total capital cost, $/kW	157.3	160.2	179.7	94.5	194.4	97.5
Total first year annual costs, mills/kWh	8.9	10.5	7.86	6.61	16.91	4.03

NA = Not available.

[a] Proposed standard of September 1978.

[b] Evaluated standards in anticipation of revision to NSPS.

[c] Promulgated standard of June 11, 1979.

[d] Evaluated standard as stringent as promulgated NSPS.

[e] Evaluated standards more stringent than promulgated NSPS.

[f] Previous standard of 1971.

TABLE 5-4. ADJUSTED CAPITAL AND ANNUAL COSTS OF OPERATIONAL LIMESTONE FGD SYSTEMS[a]

Utility name unit name	$/kW, capital	mills/kWh, annual
Alabama Electric Coop		
Tombigbee 2 and 3	35.1	2.91
Arizona Public Service		
Cholla 1	74.6	4.36
Cholla 2	148.7	7.64
Central Illinois Light		
Duck Creek 1	121.3	7.96
Indianapolis Power & Light		
Petersburg 3	148.4	8.59
Kansas City Power & Light		
LaCygne 1	81.4	6.89
South Carolina Public Service		
Winyah 2	43.1	1.80
South Mississippi Electric Power		
R.D. Morrow, SR. 1 and 2	108.7	6.01
Southern Illinois Power Coop		
Marion 4	110.8	7.12
Springfield City Utilities		
Southwest 1	133.5	7.66
Tennessee Valley Authority		
Widows Creek 8	145.1	8.56
Average	99.6	6.03

[a] The variability of these figures occurs in part because FGD systems installed on some boilers do not accommodate 100% of the boiler flue gas. The costs for such systems are proportionately lower than those for full capacity FGD systems. This is magnified by the conventional use of gross kW for the $/kW figure and net kW for the mills/kWh figure, regardless of the % of the flue gas scrubbed. These figures represent the capital and annual costs required to bring the individual units into compliance.

of a separate reheat system, the effects of more stringent SO_2 emission standards, more elaborate sludge disposal strategies, and one spare scrubber module for extra capacity.

SECTION 6

CONCLUSIONS

The discussion in the preceding sections of this paper indicates that the significant rate of growth observed in the development and application of FGD technology for coal-fired utility boilers has been matched by the considerable improvements observed in the performance of the operational systems. With respect to the latter, the most significant improvement in the performance of the operational systems involves the increased level of dependability observed for the high sulfur coal units. During the past 2 years, the dependability of these systems has improved to a level which approaches that observed for the low sulfur coal units. It is anticipated that this trend will continue and will be reflected in less startup and commercial operating problems for systems now being placed in service or planned for service.

Promoted by the requirements set forth in the Clean Air Act Amendments and the pursuant NSPS, application of FGD to all new coal-fired utility boilers constructed in the near future is anticipated.

REFERENCES

1. U.S. Department of Energy. Energy Information Administration. Office of Coal and Electric Power Statistics. Electric Power Statistics Division. Inventory of Power Plants in the United States, December 1979. Publ. No. DOE/EIA-0095(79).

2. Rittenhouse, R. C. New Generating Capacity: When, Where, and By Whom. Power Engineering 82(4):57, April 1978.

3. Bechtel National, Inc. Economic and Design Factors for Flue Gas Desulfurization Technology. EPRI CS-1428, April 1980.

4. PEDCo Environmental, Inc. Particulate and Sulfur Dioxide Emission Control Costs for Large Coal-Fired Boilers. EPA-450/3-78-007 (NTIS PB 281271), February 1978.

5. Augustine, F., S. D. Severson, and J. L. Winter. Economics of Four FGD Systems. Prepared for Electric Power Research Institute by Stearns-Roger Engineering Corporation under EPRI Research Project No. 1180-3, Draft Report, November 1979.

6. Beychok, M. R. Comparative Economics of Advanced Regenerable Flue Gas Desulfurization Processes. EPRI CS-1381, March 1980.

7. Oliver, E. D. and K. Semrau. Investigation of High SO_2 Removal Design and Economics, Volume 2: Economics. EPRI CS-1439, June 1980.

8. Tomlinson, S. V., F. M. Kennedy, F. A. Sudhoff, and R. L. Torstrick. Definitive SO_x Control Process Evaluations: Limestone, Double-Alkali, and Citrate FGD Processes. EPA-600/7-79-177 (NTIS PB 80-105828), August 1979.

APPENDIX A

FLUE GAS DESULFURIZATION INFORMATION SYSTEM

BACKGROUND

The most significant product of EPA's utility FGD survey program is a quarterly summary report generated from a computerized data base known as the Flue Gas Desulfurization Information System (FGDIS). This data base represents the latest development in this program. Previously, manually updated and semi-automated data files were used to store and retrieve information. The increased emphasis on FGD for SO_2 control (resulting from its commercial development) necessitated a more efficient data storage/retrieval system for processing and transmitting these data. In the fall of 1978, FGDIS was developed to meet this need.

DESCRIPTION

Design and performance data for both the operational and planned domestic utility FGD systems are stored in the FGDIS. Also stored are data on operational domestic scrubbers for removal of particulate matter and data on operational FGD systems applied to coal-fired utility boilers in Japan.

The design data contained in FGDIS encompass the entire emission control system and the power-generating unit to which it is applied. Descriptions include location, standards limiting emissions of SO_2 and particulate matter, power-generating capacity, boiler and stack information, average fuel analyses, and other more general data. Input of design data specific to FGD systems ranges from general information such as process type, system supplier, and initial system startup date to more specific component design information and operating parameters such as absorber type, gas and liquid flow rates, and pressure drop. Also included in the data are descriptions of the methods of solids concentrating and waste disposal, flue gas reheat, and mist elimination, and information on capital costs and annual revenue requirements of FGD systems.

For operational FGD systems, the FGDIS stores comprehensive performance data, including periodic dependability parameters and the service times (operating, forced-outage, and scheduled outage) from which they are calculated. Where available, actual system SO_2 and particulate matter removal efficiencies are included (and qualified). Problems encountered with system operation and solutions implemented to correct them are described. The performance of the FGD-equipped boiler is described in terms of service time, production (kWh), and capacity factor.

Figure A-1 presents a complete FGDIS structure diagram illustrating all of the information areas and some of the key data entries contained in the system, as well as the hierarchy associated with them in the data base. General unit data are at level 0, whereas most of the specific component data are at Level 3.

CAPABILITIES

In addition to being used to generate a quarterly report, FGDIS is also available for direct on-line access. This important function not only provides interested parties with an opportunity to examine data that are too specific for convenient inclusion in the quarterly report, but it also provides immediate access to information that has been loaded into the system but not yet published (i.e., information that has become available during the period between quarterly reports). Information is gathered, reduced, verified, and loaded into the FGDIS on a continual basis to ensure that the files remain current and complete.

Access to the FGDIS data files and manipulation of these data are accomplished via MRI System 2000®. This comprehensive data base management system offers extensive data retrieval capabilities. The set of user-oriented commands provided are flexible enough to satisfy virtually any information need. The PRINT command will produce the compilation of a simple sequential list, or a set of report writer commands will produce a tabulation of the requested data in a predetermined report format. Utilization of system functions (average, standard deviation, summation, maximum value, minimum value) will elicit statistical analyses of the numerical data in the files. In addition, the data requested through the available commands can be selectively limited by a set of criteria included in the commands. This feature facilitates examination of design or performance parameters for a specific unit or a specific process type, and so on. The retrieval possibilities are limited only by the needs and imagination of the user.

The FGDIS files are stored at EPA's National Computer Center (NCC) in Research Triangle Park, North Carolina, and are accessible via a nationwide communications network consisting of

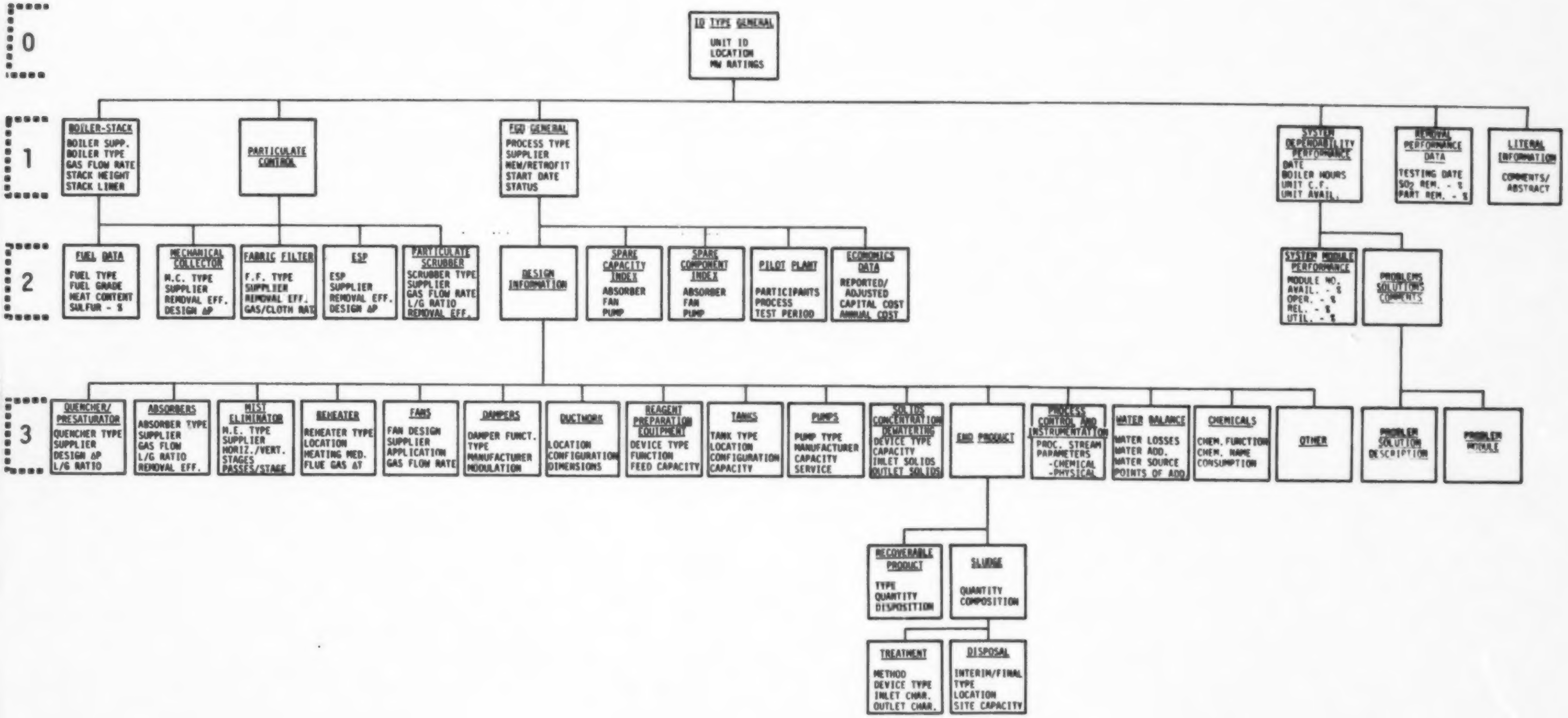

0
1
2
3
ID TYPE GENERAL
UNIT ID
LOCATION
MW RATINGS
BOILER-STACK
BOILER SUPP.
BOILER TYPE
GAS FLOW RATE
STACK HEIGHT
STACK LINER
PARTICULATE CONTROL
FGD GENERAL
PROCESS TYPE
SUPPLIER
NEW/RETROFIT
START DATE
STATUS
SYSTEM DEPENDABILITY PERFORMANCE
DATE
BOILER HOURS
UNIT C.F.
UNIT AVAIL.
REMOVAL PERFORMANCE DATA
TESTING DATE
SO2 REM. - %
PART REM. - %
LITERAL INFORMATION
COMMENTS/ ABSTRACT
FUEL DATA
FUEL TYPE
FUEL GRADE
HEAT CONTENT
SULFUR - %
MECHANICAL COLLECTOR
M.C. TYPE
SUPPLIER
REMOVAL EFF.
DESIGN ΔP
FABRIC FILTER
F.F. TYPE
SUPPLIER
REMOVAL EFF.
GAS/CLOTH RAT
ESP
ESP
SUPPLIER
REMOVAL EFF.
DESIGN ΔP
PARTICULATE SCRUBBER
SCRUBBER TYPE
SUPPLIER
GAS FLOW RATE
L/G RATIO
REMOVAL EFF.
DESIGN INFORMATION
SPARE CAPACITY INDEX
ABSORBER
FAN
PUMP
SPARE COMPONENT INDEX
ABSORBER
FAN
PUMP
PILOT PLANT
PARTICIPANTS
PROCESS
TEST PERIOD
ECONOMICS DATA
REPORTED/ ADJUSTED
CAPITAL COST
ANNUAL COST
SYSTEM MODULE PERFORMANCE
MODULE NO.
AVAIL. - %
OPER. - %
REL. - %
UTIL. - %
PROBLEMS SOLUTIONS COMMENTS
QUENCHER/ PRESATURATOR
QUENCHER TYPE
SUPPLIER
DESIGN ΔP
L/G RATIO
ABSORBERS
ABSORBER TYPE
SUPPLIER
GAS FLOW
L/G RATIO
REMOVAL EFF.
MIST ELIMINATOR
M.E. TYPE
SUPPLIER
HORIZ./VERT.
STAGES
PASSES/STAGE
REHEATER
REHEATER TYPE
LOCATION
HEATING MED.
FLUE GAS ΔT
FANS
FAN DESIGN
SUPPLIER
APPLICATION
GAS FLOW RATE
DAMPERS
DAMPER FUNCT.
TYPE
MANUFACTURER
MODULATION
DUCTWORK
LOCATION
CONFIGURATION
DIMENSIONS
REAGENT PREPARATION EQUIPMENT
DEVICE TYPE
FUNCTION
FEED CAPACITY
TANKS
TANK TYPE
LOCATION
CONFIGURATION
CAPACITY
PUMPS
PUMP TYPE
MANUFACTURER
CAPACITY
SERVICE
SOLIDS CONCENTRATION DEWATERING
DEVICE TYPE
CAPACITY
INLET SOLIDS
OUTLET SOLIDS
END PRODUCT
PROCESS CONTROL AND INSTRUMENTATION
PROC. STREAM PARAMETERS
-CHEMICAL
-PHYSICAL
WATER BALANCE
WATER LOSSES
WATER ADD.
WATER SOURCE
POINTS OF ADD.
CHEMICALS
CHEM. FUNCTION
CHEM. NAME
CONSUMPTION
OTHER
PROBLEM SOLUTION DESCRIPTION
PROBLEM MODULE
RECOVERABLE PRODUCT
TYPE
QUANTITY
DISPOSITION
SLUDGE
QUANTITY
COMPOSITION
TREATMENT
METHOD
DEVICE TYPE
INLET CHAR.
OUTLET CHAR.
DISPOSAL
INTERIM/FINAL
TYPE
LOCATION
SITE CAPACITY

Figure A-1. FGDIS structure diagram.

local telephone numbers in 21 cities and WATS services. Arrangements are currently being made so that persons interested in gaining access to the FGDIS can obtain account numbers, training, and additional information from the National Technical Information Service (NTIS), Springfield, Virginia. In addition to providing continual on-line access capability, NTIS also can process selective information requests for limited information needs that do not warrant acquisition of a permanent computer account number (single requests for specific tabulated information).

APPENDIX B

REPORTED AND ADJUSTED CAPITAL AND ANNUAL COSTS FOR OPERATIONAL FGD SYSTEMS

	Reported				Adjusted			
	Capital cost	Total/kW	Annual	Mills/kWh	Capital	$/kW	Annual	Mills/kWh
Alabama Electric								
Tombigbee 2	6,992,100	27.4	217,464	0.33	8,949,850	35.1	3,893,050	2.91
Tombigbee 3	6,992,100	27.4	217,464	0.33	8,949,850	35.1	3,893,050	2.91
Arizona Public Service								
Cholla 1	6,550,000	52.0	NA	NA	9,400,764	74.6	3,130,900	4.36
Cholla 2	44,352,000	168.0	1,003,568	0.75	39,748,800	148.7	10,221,000	7.64
Central Illinois Light								
Duck Creek 1	30,583,000	73.5	10,851,000	5.54	50,452,200	121.3	17,143,200	7.96
Central Illinois Public Service								
Newton 1	107,831,000	174.8	NA	NA	149,388,600	242.1	44,003,900	13.44
Columbus & Southern Ohio Electric								
Conesville 5	22,836,000	55.6	9,132,726	5.81	76,423,700	93.0	28,288,970	6.62
Conesville 6	22,836,000	55.6	9,132,726	5.81	76,423,700	93.0	28,288,970	6.62
Duquesne Light								
Elrama 1-4	59,541,000	116.8	21,027,000	7.18	87,852,700	172.3	30,006,600	10.82
Phillips 1-6	50,356,000	122.8	18,301,000	11.32	78,993,100	192.7	35,558,600	18.64
Indianapolis Power & Light								
Petersburg 3	55,724,000	99.5	NA	NA	78,967,000	148.4	25,189,600	8.59
Kansas City Power & Light								
Hawthorn 3	3,220,000	29.3	346,441	1.15	6,329,500	57.5	2,436,200	4.39
Hawthorn 4	3,220,000	29.3	346,441	1.15	6,329,500	57.5	2,436,200	4.39
LaCygne 1	46,900,000	53.7	7,413,047	4.99	71,124,100	81.4	32,189,700	6.89
Kentucky Utilities								
Green River 1-3	4,500,000	70.3	364,005	5.20	7,682,400	120.0	2,817,900	8.25
Louisville Gas & Electric								
Cane Run 4	12,647,000	66.6	960,301	1.29	20,045,000	105.5	5,334,000	5.15
Cane Run 5	12,481,000	62.4	763,443	0.92	17,146,000	85.7	4,975,500	4.56
Cane Run 6	20,596,900	71.5	NA	NA	23,205,000	80.6	8,867,600	5.79
Mill Creek 3	18,846,880	42.6	321,463	1.25	26,751,200	60.5	8,855,500	3.70
Paddy's Run 6	3,700,000	52.9	NA	NA	7,288,000	104.1	3,746,200	10.36
Minnkota Power Cooperative								
Milton R. Young 2	44,119,500	92.5	1,779,375	1.25	62,872,500	131.0	13,914,300	5.99
Monongahela Power								
Pleasants 1	65,693,400	106.3	9,015,879	2.73	70,058,000	113.4	26,148,300	7.92
Montana Power								
Colstrip 1	36,500,000	101.4	6,128,000	2.97	48,183,500	133.8	14,719,250	7.79
Colstrip 2	36,500,000	101.4	6,128,000	2.97	48,183,500	133.8	14,719,250	7.79
Nevada Power								
Reid Gardner 1	5,363,378	42.9	251,514	0.46	9,992,150	79.9	3,314,600	5.29
Reid Gardner 2	5,363,378	42.9	251,514	0.46	9,992,150	79.9	3,314,600	5.29
Reid Gardner 3	14,200,565	113.6	131,824	0.23	17,307,000	138.5	4,247,300	6.78
Northern Indiana Public Service								
Dean H. Mitchell 11	18,192,040	157.4	2,414,589	13.02	26,999,900	233.6	9,832,000	18.37
Northern States Power								
Sherburne 1	34,982,000	49.3	2,716,758	0.75	67,996,450	94.4	18,990,800	4.63
Sherburne 2	34,982,000	49.3	2,716,758	0.75	67,996,450	94.4	18,990,800	4.63
Pacific Power & Light								
Jim Bridger 4	49,643,000	90.3	NA	NA	59,732,500	108.6	19,440,100	6.71
Pennsylvania Power								
Bruce Mansfield 1	110,639,000	120.6	9,979,850	3.28	121,270,800	132.3	44,890,750	9.56
Bruce Mansfield 2	110,639,000	120.6	9,979,850	3.28	121,270,800	132.3	44,890,750	9.56
Philadelphia Electric								
Eddystone 1	30,856,000	285.7	3,808,000	6.37	20,206,400	187.1	6,296,400	10.53
Public Service Company of New Mexico								
San Juan 1	47,944,410	132.8	NA	NA	92,034,400	254.9	31,930,100	17.86
San Juan 2	47,985,000	137.1	NA	NA	90,608,200	258.9	31,483,100	18.07
South Carolina Public Service Authority								
Winyah 2	6,646,000	23.7	527,000	0.29	12,060,300	43.1	2,648,100	1.80
South Mississippi Electric								
R.D. Morrow 1	10,896,000	53.7	NA	NA	22,056,750	108.7	6,162,250	6.01
R.D. Morrow 2	10,896,000	53.7	NA	NA	22,056,750	108.7	6,162,250	6.01
Southern Illinois Power Coop								
Marion 4	15,200,930	87.9	859,453	1.03	19,177,750	110.8	6,525,600	7.12
Southern Indiana Gas & Electric								
A.B. Brown 1	12,495,000	47.2	1,850,565	1.30	21,477,900	81.1	7,252,100	5.10
Springfield City Utilities								
Southwest 1	16,744,500	86.3	778,749	1.20	25,904,900	133.5	7,413,800	7.66
Tennessee Valley Authority								
Widows Creek 8	47,900,000	87.1	14,576,400	7.80	79,785,300	145.1	25,140,300	8.56
Utah Power & Light								
Hunter 1	24,400,000	56.7	NA	NA	29,625,000	68.9	9,492,200	4.17
Huntington 1	27,090,000	63.0	2,946,400	1.27	35,498,200	82.6	12,033,100	5.28

NA = Not available.

The Department of Energy's

Flue Gas Desulfurization

Research and Development Program

Edward C. Trexler, P. E.
U. S. Department of Energy
Office of Coal Utilization

The Department of Energy's flue gas desulfurization (FGD) research and development activities are conducted as part of the Advanced Environmental Control Technology Program (AECT) which is managed within the organization of the Assistant Secretary for Fossil Energy. This new AECT program was initiated in FY 1979 with a goal to identify, research, develop, refine and demonstrate cleanup equipment that will clean flue gas for compliance with existing and anticipated environmental pollution regulations, and equipment that will remove the undesirable components from coal derived gas streams to assure reasonable life for utilization equipment such as gas turbines and fuel cells. The flue gas cleanup portion of the AECT program budget amounted to $2.7 million in FY 1979 and $20.1 million in FY 1980.

The FGD project is divided into two parallel efforts identified by the scheduled completion dates as very near-term (end 1983) and near-term (end 1986). The very near-term effort aims at improving the SO_2 removal efficiency and reducing the waste disposal problems of conventional lime/limestone scrubbers. This is being done in coordination with EPRI and EPA scrubber improvement programs, through private sector scrubber instrumentation and analysis, by tests at TVA and other utility prototype and full-scale scrubber facilities, and by transfer of process improvement information. The near-term effort is aimed at supporting newer technology SO_2 removal processes that include non-regenerable (throwaway) and regenerable systems that produce potentially marketable by-products such as sulfur and sulfuric acid. These technologies are, or will, be under experimental test at Fossil Energy Technology Centers, under prototype testing by DOE and EPRI at TVA and other sites, and under initial commercial use evaluation by DOE at power stations and industrial plants. As these technologies mature, private industry will be encouraged to cost-share development with the Government. Information on progress will be disseminated via reports, symposia, plant visitations, demonstrations and workshops.

The Department of Energy's
Flue Gas Desulfurization
Research and Development Program

NATIONAL PRIORITIES IN ENERGY AND ENVIRONMENT

The Nation's entrance into the 1980's is characterized by the need to solve difficult and interrelated problems. High on this priority list are the needs to significantly reduce oil imports, to protect and enhance the environment, and to improve the economic posture of the Nation through increased national productivity. That these needs are important to the Nation is evidenced by the abundance of contemporary legislative activity which promotes both the diminished use of oil and gas through coal utilization and the enhancement of the environment. Explicit in these legislative acts is the need for achieving these goals within the bounds of economic constraint. Meeting these goals will require the coordinated effort of both the private and public sector. This paper seeks to promote such coordinated effort by presenting the Department of Energy's Flue Gas Desulfurization Research and Development Plans. It is our desire that this summary serve as a focal point for new and improved communication, and that the end result will be success through a better coordinated effort.

The approach in this paper is to identify the energy challenge in terms of flue gas desulfurization system needs, to introduce you to our new cleanup technology efforts and how the FGD program is oriented to other DOE programs, to note our special relationships with EPA, TVA and EPRI, and to discuss in some detail particular programs which we are pursuing. In addition, we would also have you join with us in examining the challenges and opportunities of the future.

THE ENERGY CHALLENGE

The oil importation reduction challenge perhaps can be best appreciated by observing our recent energy flow from supply through consumption, and by comparing consumption with domestic supply. Domestic and imported supply[1] in 1977 was:

Supply	Quads/Yr.
Domestic Coal	15.9
Domestic Natural Gas	22.7
Domestic Oil	16.68
Imported Oil	18.91

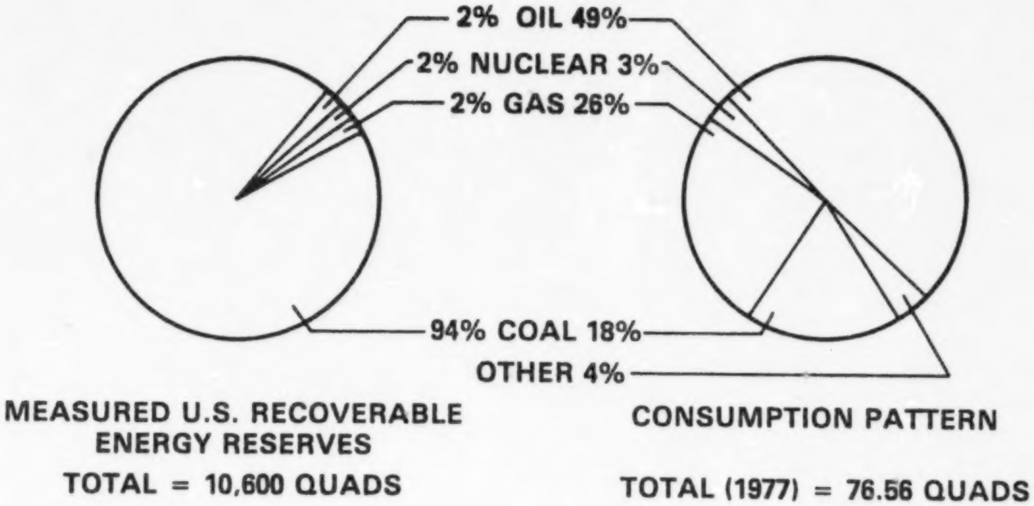
Consumption[1]/ in key sectors in 1977 was:

Sector	Consumption (Quads/Yr.) Coal	N. Gas	Oil
Electric Energy Generation	10.64	3.26	3.45
Residential/Commercial	.22	7.21	5.99
Industrial	3.14	8.65	7.60
Transportation	0.0	.54	20.0

A comparison on a percentage basis between domestic reserves and consumption is given by Figure 1.

Figure 1

U.S. RESERVES VS. U.S. CONSUMPTION



Clearly, it can be seen that the Nation needs a substantial shift in consumption from oil and gas which are not abundant, to coal which is. Because of the nature of the respective markets, it would appear easier to accomplish this shift initially from the more centralized consumers such as the utilities and the major industrial plants. The administration has set as a goal that the oil and gas consumption of this sector be reduced fifty percent (50%) of present consumption by 1990 and legislation has been enacted accordingly.

The interrelationship of our energy challenge with environmental goals was previously noted. In the near-term, we must burn more coal and we must burn it cleanly and economically, and this means we need additional FGD options. Key environmental regulations affecting coal utilization are outlined in Figure 2.

Figure 2

ENVIRONMENTAL REGULATIONS
AFFECTING COAL UTILIZATION

Clean Air Act - 1977
- National Ambient Air Quality Standards
- New Source Performance Standards

  - Prevention of Significant Deterioration Regulations
  - Nonattianment Policy
  - State Implementation Plans

Resource Conservation and Recovery Act - 1976
Toxic Substances and Control Act - 1976
Clean Water Act - 1977
Safe Drinking Water Act - 1974

Further, it is to be noted that the acid rain phenomena has been receiving considerable attention recently. This interest could result in new legislation and the need for retrofitting a new breed of low cost FGD systems into many existing coal burning installations if such sources are proven to be major contributors to the problem.

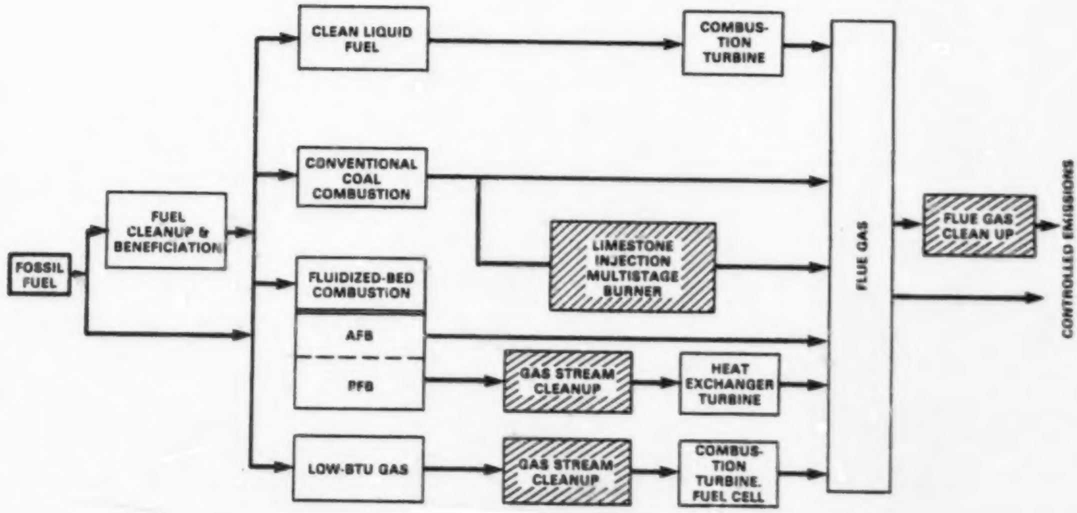
In summary, in terms of R&D objectives, we need the early supply of an assortment of systems which enable utilities and major industrial users to operate reliably and economically on coal or coal derived fuels while meeting all present and anticipated environmental regulations. Further, it is important that some of these systems be particulary oriented toward retrofit applications.

ORIENTING THE DOE FGD ACTIVITY AND PROGRAM

The Department of Energy, Fossil Energy Assistant Secretary, pursues these R&D goals with a broad based program of which the Flue Gas Desulfurization Program is a part. The DOE program is basically a private sector assistance program. The Department seeks to identify technologies with high potential public benefit and seeks to promote their accelerated development and demonstration by assuming some of the financial burden and risk. The orientation of the FGD program to certain other FE programs can be seen by Figure 3.

Figure 3

CLEANUP TECHNOLOGY CONTROL OPTIONS

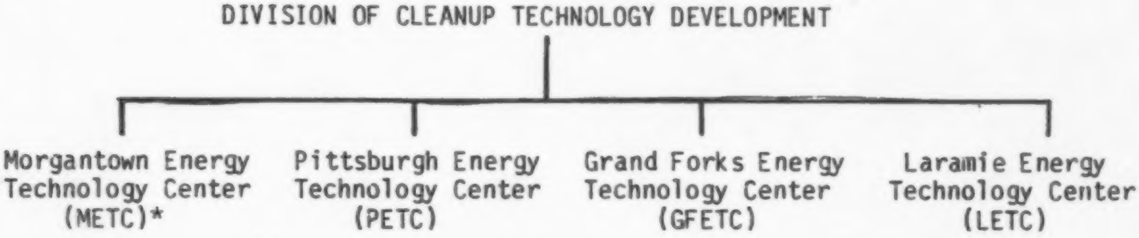

CLEAN LIQUID FUEL
COMBUSTION TURBINE
CONVENTIONAL COAL COMBUSTION
FUEL CLEANUP & BENEFICIATION
FOSSIL FUEL
LIMESTONE INJECTION MULTISTAGE BURNER
FLUIDIZED-BED COMBUSTION
AFB
PFB
GAS STREAM CLEANUP
HEAT EXCHANGER TURBINE
LOW-BTU GAS
GAS STREAM CLEANUP
COMBUSTION TURBINE, FUEL CELL
FLUE GAS
FLUE GAS CLEAN UP
CONTROLLED EMISSIONS

The Flue Gas Desulfurization Program is operated from the Office of Coal Utilization's Division of Cleanup Technology Development. Other programs operated from the Division are crosshatched in the Figure. Cleanup technology development is pursued through DOE Field Technology Centers as shown below by Figure 4.

Figure 4

CLEANUP TECHNOLOGY DEVELOPMENT

DEPARTMENT OF ENERGFY
ASSISTANT SECRETARY FOR FOSSIL ENERGY
OFFICE OF COAL UTILIZATION


DIVISION OF CLEANUP TECHNOLOGY DEVELOPMENT
Morgantown Energy Technology Center (METC)*
Pittsburgh Energy Technology Center (PETC)
Grand Forks Energy Technology Center (GFETC)
Laramie Energy Technology Center (LETC)

*Lead Center

The Cleanup Technology Division has sought, since its creation in 1979, to build on the excellent FGD technology foundation layed down by the private sector and by EPA, TVA and EPRI. I am personally grateful for the many reports from them which have afforded us the opportunity to understand and assess the technological choices. Much of our initial effort has been in providing support to programs initiated by these organizations, and we intend to continue this approach along with our modest in-house efforts, and to significantly expand our joint efforts with the private sector. The importance seen for this program within DOE is evidenced by its growth from a modest $2.7M in FY 1979 to a requested $21.0M in FY 1981. Our FY 1982 request maintains the momentum of this rapidly growing effort. We believe we will contribute by bringing the energy perspective into FGD development.

DOE FGD R&D PROGRAM

Although the DOE FGD Program includes some effort aimed at improving the reliability, operability and performance of conventional lime/limestone scrubbers, and includes some attention to new FGD approaches, the majority of our effort is going into what might be called the emerging or advanced FGD systems. This affords us the opportunity to select and pursue those particular efforts which would appear to offer the mose benefits for the markets which need to be served in order that coal utilization can be maximized in the shortest time.

Figure 5 describes those technologies which we have tentatively chosen to evaluate and how and when these evaluations might lead to large scale utility and industrial demonstration.

Figure 5

EMERGING FGD SYSTEMS

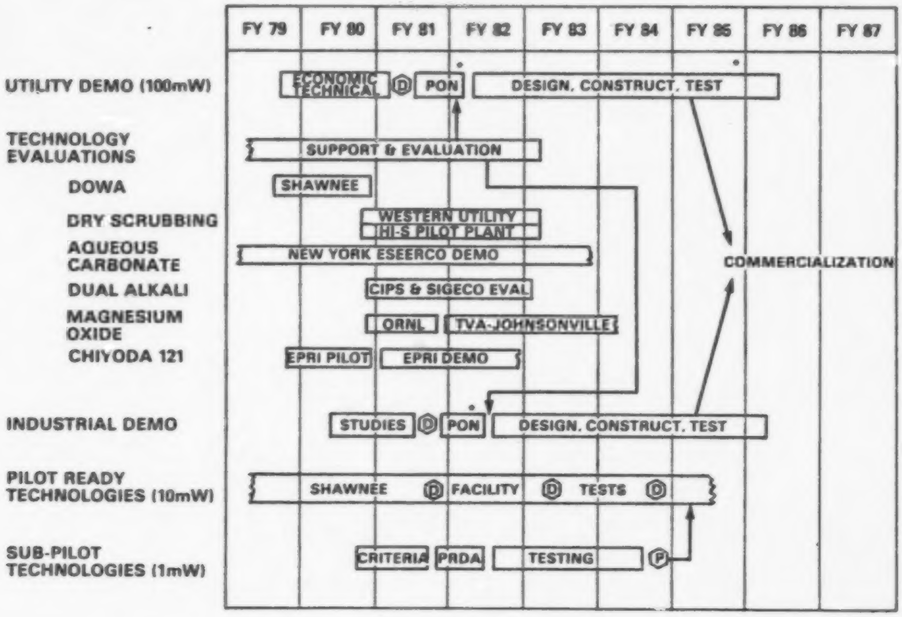


Figure 6 identifies our key near-term needs and the candidates which we are evaluating to meet these needs.

Figure 6

KEY NEAR-TERM NEEDS/CANDIDATES

Need	Candidates
o Reliable, low cost, retrofitable FGD systems for eastern coal applications which produce an easily disposable dry or gypsum waste product.	o Forced Oxidation Systems o Systems such as: - Chiyoda 121 - DOWA o Spray Dryers
o Reliable, low cost FGD systems for western coal applications with low water consumption and manageable waste products	o Dry Injection o Spray Dryers

- Reliable, low cost regenerable systems which utilize coal for reduction or regeneration

- PETC technology data base for:
  - Improved steam stripping
  - Direct coal reduction
  - Improved copper oxide/ gasifier system

Detailed discussions of these needs and the primary candidates which are being evaluated are included in the paragraphs which follow.

EASTERN NEEDS

Foremost on our list of key needs is the need to provide by the early 1980's reliable, low cost, retrofitable FGD systems for eastern coal applications, which produce easily disposal waste products such as gypsum or dry insoluable solids rather than sludge. The candidates for this need would appear to include the newer forced oxidation systems, systems such as CHIYODA 121 and DOWA and spray dryer systems. We are particularly encouraged by the recently reported improved stoichiometrics for spray dryer processes which show them to have economic advantages even with higher sulfur coal. Our evaluation programs are as follows:

- EVALUATION OF FORCED OXIDATION SYSTEMS

  DOE will study data from recent full-scale commercial forced oxidation systems and compare them with the projected qualities of CHIYODA 121 and DOWA.

We expect to complete this study in January and the results might lead us to initiate an evaluation effort.

- EVALUATION OF GYPSUM WASTE SYSTEMS

  - DOE has tentative plans to join with EPRI in evaluating a full size CHIYODA 121 module.

  - DOE may support additional DOWA efforts at the TVA Shawnee Test Facility.

We also believe that much can be learned by carefully studying the results of the recently completed CHIYODA 121 pilot scale (23 MWe) tests.

- EASTERN COAL SPRAY DRYER EVALUATIONS

  - Eastern Coal Spray Dryer development/evaluation at pilot scale (RFP - early FY 1981 award)

  - Spray Dryer evaluation at PETC 500 #/hr coal-fired boiler.

  - Spray Dryer performance characterization at ANL on 170,000 #/hr steam boiler (Preliminary)

- Joint EPA/DOE/EPRI Spray Dryer characterization at 100 MWe utility unit (Preliminary)

As noted previously, the optimistic projection for the application of spray dryers to eastern coals is recent and, accordingly, our program to increase emphasis in this area is not completely in place.

Our primary approach is to pursue this evaluation through the private sector and, accordingly, we have been preparing an RFP for such an evaluation. This RFP, which is scheduled for release in October, offers to fund pilot scale testing of eastern coal optimized spray dryer on a slate of eastern coals, and offers further to fund the conceptual design and economic evaluation of commercial scale units.

Parallel with this effort, we propose to obtain parametric performance data on a subpilot unit at our Pittsburgh Energy Technology Center (PETC), and to take advantage of the installation of a spray dryer being installed on a 170,000 Lb/Hr. steam boiler firing eastern coal at the Argonne National Laboratory (ANL).

Further, it is to be noted that DOE, in conjunction with EPA and EPRI, have been discussing Spray Dryer characterization testing on a 100 MWe utility unit and such a unit could be used to verify, at a large scale, the performance projections derived from pilot scale evaluations.

WESTERN NEEDS

For western markets, we see the need for reliable, low cost systems, with low water consumption and manageable waste products. Key facets of our western applications program are as follows:

- o EVALUATION OF DRY SCRUBBERS FOR WESTERN APPLICATIONS (GFETC)
    - Field testing of full-scale utility Spray Dryers with lime and sodium reactants.
    - Continued testing and evaluation of dry injection of alkaline ash, nahcolite and trona and the regeneration of reactants.

REGENERATION WITH COAL

We see the need for reliable, low cost regenerable systems which can utilize coal for reduction or regeneration, and we are approaching this need at this time with in-house laboratory tests and studies. This program is as follows:

- o PETC TECHNOLOGY BASE FOR IMPROVED REGENERABLE FGD SYSTEMS
    - Model the reaction dynamics for direct reduction of SO_2 with coal. Verify at bench scale.
    - Measure SO_2 partial pressures for prospective organic absorbants to optimize absorption/steam stripping systems.

- Evaluate at PDU scale a fluid bed reactor copper oxide system employing coal gas as a reducing agent.

It is our desire to produce a data base from which the private sector might create or optimize improved regnerable systems which use coal instead of oil or gas for reduction or regeneration.

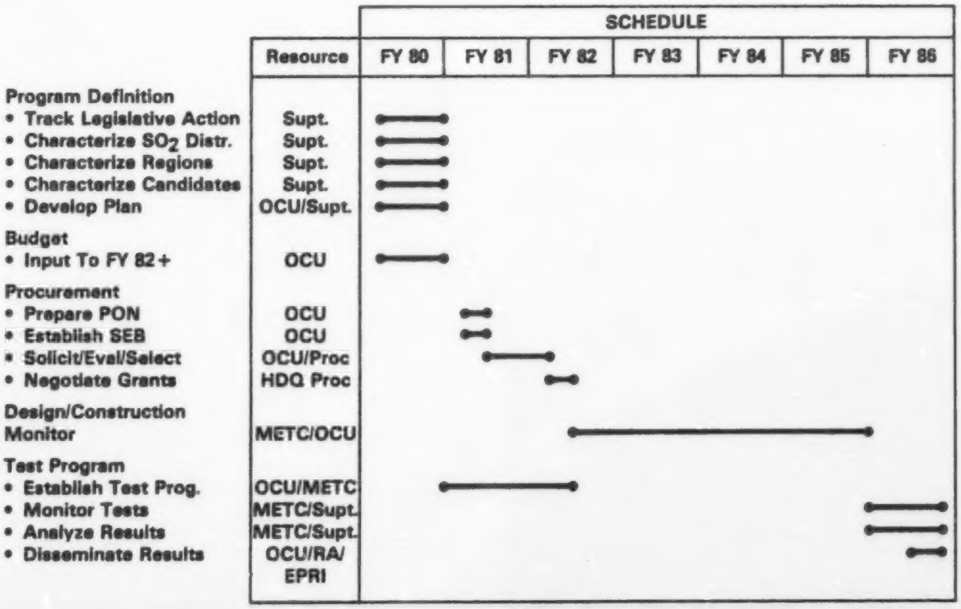
POTENTIAL NEW CHALLENGES/OPPORTUNITIES

As noted previously, the attention being given to existing pollution problems such as acid rain could precipitate the development of a new breed of low cost, retrofitable, less than NSPS capture systems to which our present plans are not addressed. We are looking carefully at the work being sponsored by EPA in this area, and we will be joining them in the Limestone Injection Multistage Burner (LIMB) effort. In addition, we have been evaluating burners, such as the staged slagging combustors, under other FE programs, which might employ limestone injection and which might lead to workable systems for such applications.

A related challenge might come from the proposed Powerplant Fuels Conservations Act of 1980 (S. 2470). While the major thrust of this proposed legislation is to mandate the conversion from oil to coal of approximately 18,000 MWe of powerplants primarily along the eastern coast, it also contains a very important "offset" provision. The offset provision seeks to offset the approximate 110,000 tons/hr of SO_2 additional emission caused by the conversion to coal, by funding the addition of advanced SO_2 removal systems to approximately 3,000 MWe of existing coal-fired units. To DOE, this is both a challenge in terms of being able to make wise choices as to appropriate systems by late 1982, and an opportunity for increased development and demonstration at a large scale. Our tentative plan for implementing the offset provision is shown below in Figure 7.

Figure 7

GENERAL PLAN



SUMMARY

In summary, DOE looks forward to joining with the flue gas cleanup community in pursuing jointly both our energy and environmental goals, and to contribute to the overall success through our perspective of the nations energy needs.

We are pleased with the opportunity to share with you our plans and our thinking, and we look forward to the opportunity to get to know all of the participants better, and to work together toward these important national goals.

EPRI RESEARCH RESULTS IN FGD: 1979 - 1980

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ABSTRACT

EPRI has a research effort of approximately $10M/year in flue gas desulfurization covering engineering evaluations, field testing, bench testing, pilot plants, prototypes and demonstrations. This paper reports selected results from projects on FGD water integration, gypsum crystallization, limestone dissolution, wet stack operation, sulfur production via RESOX, absorption/steam stripping, cyclic reheat, and integrated emission control. A brief review of current demonstration plans and program emphasis is also included.

EPRI RESEARCH RESULTS IN FGD
1979 - 1980

EPRI'S WORK IN FGD

The Electric Power Research Institute (EPRI), as the research arm of the U.S. electric utility industry, has established a research and development program in flue gas desulfurization. In this area, the Institute will fund approximately ten million dollars of R&D work each year over the next five years. Projects include engineering evaluations, laboratory testing, pilot plant work, prototype development, demonstration installations and field testing.

CONTENTS OF THIS PAPER

In this paper are presented recent data from selected EPRI projects in the areas of FGD field testing, economic evaluations, limestone dissolution, wet stacks operation, FGD water integration, cyclic reheat, crystallization, sulfur production via RESOX, absorption/stream stripping, and integrated emission control. Also included in the paper is a discussion of EPRI's R&D program emphasis in the next few years. Each project that has significant recent results is discussed under a separate heading for that project.

OBJECTIVES OF THE PROGRAM

EPRI's FGD research efforts are designed to meet one or more of the following objectives:

Reduce costs:	Reduce capital, operating, maintenance and disposal costs.
Improve reliability:	Identify reliable systems or components; develop improved materials; identify mechanisms and modes of failure, and repair requirements.
Improve resource utilization:	Improve energy efficiency; reduce dependence on oil, electricity and gas; reduce water consumption and discharges; improve by-product utilization.

The specific projects discussed in this paper represent recent EPRI funded work not reported separately at this symposium. Papers are being presented in other sessions covering successful testing of a 23 MW Chiyoda Thoroughbred 121 system with gypsum stacking at Gulf Power's plant Scholz, joint EPRI/TVA/UOP testing of the 10 MW Dowa prototype, and EPRI solid waste disposal efforts. These topics will not be covered further in this paper.

Projects discussed in this paper:

Subject	EPRI Project Number	Project Description	Contractors	EPRI Contact
Absorption Steam Stripping	RP1402-2 RP1258-1 -2,3,4	Lab and pilot development of Flakt Boliden citrate and novel steam stripping processes	U of Texas at Austin (Dr. Rochelle), TVA, Stearns-Roger, Radian	D. A. Stewart
Characterization of FGD	RP1410-3	Test two FGD units comprehensively to establish operating capabilities and material and energy balances.	Black & Veatch, MRI, PEDCO, TRW	R. G. Rhudy
Chemistry	TPS79-747 RP982-21	Mg dissolution from limestone to improve scrubber performance.	Radian Corp.	D. A. Stewart
Corrosion	RP982-14 RP982-19	Lab testing of corrosion and erosion in FGD.	Battelle Columbus SumX Corp.	R. G. Rhudy C. E. Dene
Crystallization	RP1031-2,3	Bench Scale sulfate crystallization	U. of Arizona (Dr. Randolph)	D. A. Stewart
Cyclic Reheat	RP1652-1	Economic and field evaluation of the cyclic reheat concept (using inlet heat to reheat)	Bechtel National Corp.	R. G. Rhudy
Economics	TPS78-760 TPS78-767	High SO_2 removal Design and Economics Vol 1 Design Vol. 2 Economics	(Companion studies) Radian Corp. SRI International	R. G. Rhudy
	RP1180-9	Economic and Design Factors for FGD Technology	Bechtel National	C. R. McGowin
Integrated Emission	RP1646-1 1870-2	Build and test pilot 2-1/2 MW integrated facility	Stearns-Roger, et al	D. V. Giovanni T. M. Morasky
SO_2 Reduction	RP784-2 RP1257-1	RESOX pilot and prototype development.	Foster Wheeler Energy Corp., et al	T. M. Morasky
Water Integration	TPS80-730	Material balance to show effect of different water sources on various FGD systems	Radian Corp.	D. A. Stewart R. Kosage
Wet Stack Designs	RP1653-1	Entrainment and engineering for wet stacks	Dynatech R/D Co.	C. E. Dene

RESULTS SUMMARY HIGHLIGHTS

Absorption/Steam Stripping	Laboratory work has confirmed the Flakt-Boliden citrate process data, identified several potential stream stripping process improvements, and set the stage for 1 MW pilot plant testing at TVA's Colbert plant facility.
Characterization	Two FGD units have been tested, the Colstrip Unit 2 of Montana Power Co. and the Conesville Unit 5 of Columbus and Southern Ohio Electric Co. Some details of the test results are given in the attached writeup.
Chemistry	Certain magnesium-containing limestones may be more reactive than high-calcium stones depending on the minerology. Three promising stones have been identified for further screening.
Corrosion	Surveyed installations and manufacturers and identified downstream ductwork, stacks, dampers and expansion joints as special problem areas. Evaluated chemical additives as corrosion inhibitors and identified N-lauroylsarcosine for further evaluation.
Crystallization	Developed calcium sulfate crystal growth predictive equations, evaluated certain crystal habit modifiers, and found a crystallizer configurations which may help in controlling gypsum crystal size.

Cyclic Reheat	The use of inlet heat to reheat exhaust gases appears to break even economically with steam reheat for moderate steam costs when using high-cost alloy for the cyclic reheat system. The low sulfur coal full-scale cyclic reheat system was tested and found to be working well. High sulfur coal cyclic reheat economics depend on construction material, degrees of reheat, and inlet flue gas temperature.
Economics	Several special purpose evaluations were performed. Regenerable processes generally are more expensive for the specific cases studied. High SO_2 removal design studies (TPS 78-760/1767) identified potential for effect of Mg in reducing high SO_2 removal costs in conventional FGD. Generalized case studies (RP1180-9) identified spray drying as a cost saving technique for western FGD and CT-121 as having low lifetime costs. Under RP784-1, the possible benefits of absorption/steam stripping combined with RESOX were identified (though these were not verified in later work).
SO_2 Reduction	Pilot work at 1 MW scale has verified RESOX suitability for different types of coals and for different SO_2 feed stream concentrations. German 42 MW prototype efforts have not shown high sulfur yields or sustained operating times.
Water Integration	Over forty material balance cases have been evaluated. Several cases show increased scrubber scaling potential with certain sources of water.

Wet Stack Design	Based on a literature search and theoretical calculations a significant portion of the water present in the stack appears due to carryover from mist eliminators. Design criteria from existing wet stacks and the problems encountered are identified in the attached write-up.

ABSORPTION/STREAM STRIPPING PROCESS
RP1258 and RP982-20

OBJECTIVE

Regenerable processes for SO_2 removal from stack gases are being investigated in order to hasten development of an economically feasible FGD process alternative to the throwaway systems. Initial cost comparisons of several processes indicated that absorption/steam stripping may be an economically competitive FGD process.

PROJECT DESCRIPTION

A project was initiated in 1978 to study the Flakt-Boliden absorption/steam stripping process as it was the most technically advanced. Laboratory confirmation of basic process and pilot plant construction were conducted concurrently followed by pilot plant tests to obtain firm data for design and cost studies.

RESULTS

In the absorption/steam stripping process, SO_2 is absorbed in a buffered solution and then stripped from the solution with steam. The stability of the dissolved SO_2 in the buffered solution is an important factor in SO_2 recovery. Loss of the dissolved SO_2 may result by disproportionation or by reaction with another component, such as the buffer or oxygen. The results of a study of the stability of SO_2 in the two most important absorbents, sodium citrate and diethylenetriamine (DETA), are shown in Table 1, along with the stripping steam requirement.

TABLE 1 Loss of SO_2 from Buffer Solutions

Buffer	Buffer Conc., M	Initial dissolved SO_2 Conc., M	Temp, °C	Loss Rate x 10^3, M/hr	Estimated Stripping Steam Rate, Kg/Kg SO_2
Citrate	0.5	0.2	140	1.2	40
	0.5	0.2	150	3.3	
	0.5	0.2	158	8.9	
	0.75	0.2	163	23.0	
DETA	2.0	0.08	139	2.9	20
	2.0	0.2	145	6.2	
	2.0	0.2	155	12.0	

DETA solutions do not appear to retain SO_2 as easily as citrate solutions. The savings in steam costs are the primary reason for continued investigation of DETA. Comparisons of stability of the absorbents, citrate and DETA, are being made.

FUTURE WORK

The pilot plant study of the Flakt-Boliden process (citrate absorbent) is currently underway at the Colbert Station of the Tennessee Valley Authority. Following analysis of the data from this test program, further pilot tests will be conducted with either citrate or DETA. The extent of the test program with DETA depends on the results of the laboratory work on DETA stability.

CHARACTERIZATIONS OF FULL-SCALE SCRUBBERS

RP1410-3

OBJECTIVE

The full-scale scrubber characterizations project was initiated in order to provide the needed data base to enable the utilities to optimize their existing systems effectively, to aid them in selecting new systems, and to provide informed utility responses to possible new emission standards.

PROJECT DESCRIPTION

The project is directed at performing extensive and detailed characterizations of the capabilities of selected, representative, currently operating, full-scale lime and limestone wet scrubbing systems. The program characterizes the performance of the selected scrubber system with respect to the following:

- Meeting emission standards and performance guarantees, with emphasis on sulfur dioxide removal.
- Quality and quantity of selected unregulated discharges for such species as organic compounds, volatile metals, fine particulates, and trace elements.
- Actual costs compared to estimated costs, including both capital and operating costs.
- Reliability, availability, and operability.

The initial scrubber systems selected for characterization are Columbus and Southern Ohio Electric Company's Conesville Unit 5 and Montana Power Company's Colstrip Unit 2, burning high sulfur eastern coal and low sulfur western coal, respectively.

RESULTS AND CONCLUSIONS

Work has been completed at Conesville and a draft final report is in review. Field testing has been completed at Colstrip and the data are being analyzed prior to preparation of a final report.

Conesville Unit 5 is a 411 megawatt power plant which began operation in January, 1977. The flue gas cleaning system for this unit consists of a cold-side electrostatic precipitator followed by a turbulent contact absorber scrubbing system capable of greater than 90 percent sulfur dioxide removal from a high sulfur coal. Magnesium containing lime is used as the scrubber additive and the scrubber sludge is stabilized by a commercial "fixation" process and stored onsite in a landfill operation.

With respect to regulated emissions, greater than 95 percent SO_2 removal was measured across one module of the two-module system. Although the net SO_2 removal is decreased for Unit 5 because of a system bypass and these measurements were short term (8 hours), the presence of a high level of dissolved alkalinity provided by the magnesium in the lime would allow a reduction of one third in pumping power (3 pumps to 2 pumps) with only a 1 to 2 percent change in the SO_2 removal.

The particulate removal across the module measured was always positive. The particulate removals may not be representative because the inlet values were higher than expected (suggesting either high inlet ESP loadings or non-optimum ESP operation) and the outlet values may be affected by SO_3 condensation across the scrubber. However, no evidence was found to indicate a significant scrubber related particulate emission increase. Removal of NO_x was insignificant.

The condensation of SO_3 across the scrubber created problems in the particulate size distribution measurements. The only particulate penetration measured, in the 0.1 to 0.2 μm range, was attributed to sulfuric acid condensation based on the size, appearance, and elemental composition of the material captured. The trace element data is still being reviewed and it is too early to present the results. The measurements of organic emissions indicated few were present and what was measured was well below its toxic level.

Average availability of the Conesville Unit 5B scrubber module from January 1979 through August 1979 was 39.2 percent. If major outages which resulted from labor problems, failure of major equipment components, and design changes are subtracted for this period, maximum expected availability would increase to about 68 percent.

The remaining 32 percent of module unavailability is due to a variety of maintenance requirements, such as cleaning plugged lines, cleaning scrubber modules, and repairing equipment which had malfunctioned. A vigorous record keeping plan has been initiated by the operating utility which will allow identification of individual maintenance problems in the future. Maintenance levels on the unit have been substantially increased and the current availability of the unit is close to the boiler availability.

CHEMISTRY OF MAGNESIAN LIMESTONE DISSOLUTION

RP982-21 and TPS80-730

OBJECTIVE

The presence of dissolved magnesium in lime or limestone FGD systems generally improves the SO_2 absorption due to an associated increase in disolved alkali. However, the magnesium present naturally in limestone is usually in the form of dolomite, which is too slowly soluble to significantly increase the magnesium ion concentration. Recently, a few limestones containing greater than 1% $MgCO_3$ (magnesian limestone) have been tested which appear to have a portion of the magnesium in soluble form.

To determine if magnesian limestones containing soluble magnesium compounds are common, a survey of the literature was conducted to locate limestone formations containing greater than 3% $MgCO_3$ but less than pure dolomite (46% $MgCO_3$). These formations have been sampled for chemical and mineralogical analyses and solubility and rate of dissolution determinations.

PROCEDURE

Samples of 12 different magnesian limestones have been taken directly from the quarries. These quarries are mostly in the east and midwest. Samples of some western U.S. limestones are also available for study. Characterization of the stones includes chemical analyses for major constituents, X-Ray diffraction to determine mineral content, and optical and electron microscopy to determine grain size. Selected stones were tested for equilibrium solubility in water by mixing a ground sample with water, agitating at a constant temperature, and analyzing with time to a constant composition. The rate of dissolution of ground limestone is determined by adding limestone to simulated FGD liquors and analyzing with time. The effects of limestone particle size, temperature, rate of agitation, pH, and initial solution composition on solution rate are being studied.

RESULTS AND CONCLUSIONS

Ten magnesian limestone samples from different formations have been subjected to X-Ray diffraction analysis. Of these stones, three contain portions of the magnesium in a form other than dolomite. A comparison of the rate of dissolution if $MgCO_3$ from these three samples will be compared to rates of a dolomite stone and calcite stone. Preliminary results from a study of the effects of variables such as particle size, temperature, rate of agitation, pH and solution concentrations on the rate indicate that these variables affect the rate of solution by different degrees for the different stones. For example, increasing temperature from 50°C to 60°C increases the rate of solution of $CaCO_3$ from Fredonia limestone but has little effect on rate from Maysville Limestone (a magnesian limestone).

FUTURE WORK

The experimental procedures described here will be used on additional limestone samples to determine if variables studied have any major effects on rate of solution of either magnesium or calcium compounds in the limestone. If the effects on solubility are not the same for each limestone, further characterization of the limestone properties will be made in an attempt to correlate limestone variables with differences in solubility behavior.

CORROSION INHIBITORS
RP982-17

OBJECTIVE

Corrosion complicated by erosion has been a continuing problem in wet scrubbing systems for flue gas desulfurization (FGD). The attempts to solve these problems have been made using coatings, linings, and various metal alloys. Despite these attempts, maintenance and replacement coats have remained very high.

Techniques such as the use of corrosion inhibitors have not been seriously investigated for corrosion prevention in FGD systems. SumX Corporation has undertaken a study for EPRI designed to determine the feasibility of using the absorption type corrosion inhibitors in lime or limestone scrubbing solutions.

The major objective of this study is to determine if absorption inhibitors can be used to lessen corrosion in FGD equipment.

PROJECT DESCRIPTION:

The work consists of laboratory experiments using electrochemical techniques to detect changes in the corrosion potential of the metal in scrubber liquors.

Data from literature as well as recommendations from inhibitor suppliers were used to select inhibitors for preliminary screening. The effect of these inhibitors on the corrosion of mild steel, 304L stainless steel and 316L stainless steel under one set of solution conditions was measured.

RESULTS AND CONCLUSIONS:

To date 10 compounds have been tested with mild steel. N-Lauroylsarcosine has shown the best inhibitor properties. Sulfite concentrations appear to have a major influence on corrosion. The formation of a reaction film can be critical to the corrosion rate.

Tests conducted with stainless steels are incomplete; however, uniform corrosion rates are very low. The major influence on corrosion observed thus far has been temperature.

Future work on this project will involve compounds related to N-Lauroylsarcosine, completing tests with 304L and 316L stainless steels. In addition, tests to determine sensitivity to inhibitor concentration and other solution characteristics will be conducted. Coupon tests with slurry solutions will be performed for extended periods with the most promising inhibitor compounds.

CORROSION AND MATERIALS OF CONSTRUCTION
RP982-14

OBJECTIVE

The objective of the materials of construction in wet scrubbers project was to comprehensively document and analyze the utility experience with materials of construction in full-scale lime and limestone wet FGD systems on boilers burning eastern or western coals. The result will be a summary of materials experience.

PROJECT DESCRIPTION

Information on field performance of construction materials was collected primarily by site visits, but also by telephone and letter contacts with FGD system operators and equipment vendors, and by literature searches. Information was collected for the following FGD system components: prescrubbers, absorbers, spray nozzles, mist eliminators, reheaters, fans, ducts, expansion joints, dampers, stacks, storage silos, ball mills, slakers, pumps, piping valves, tanks, thickeners, agitators, rakes, vacuum filters, centrifuges, and pond linings.

Materials documentation and analysis include successes, failures, reasons for success or failure, failure mechanisms, and relative costs of various materials. Detailed trip reports on each site visit are included in an appendix. The results are designed to be a first step in aiding utilities and FGD equipment suppliers in selecting materials that will perform satisfactorily at minimum expense.

RESULTS AND CONCLUSIONS

Stack linings and outlet ducts (beyond outlet dampers) are the scrubber components that have a significant history of materials problems and are critical components in that failures may require complete boiler shutdown and loss of generating capacity for lengthy periods due to the lack of standby components or bypass capability.

The performance of a stack lining depends on whether the scrubbed gas is delivered to the stack wet or reheated, and whether or not the stack is also used for hot bypassed gas. These factors appear to have a strong effect on the performance of lining materials, in spite of differences in fuel sulfur, application techniques (e.g., surface preparation or priming), operating procedures (e.g., thermal shock), design (e.g.,annulus pressurization), and other factors which can affect performance.

Inlet and bypass ducts are generally not a major problem area for utilities with scrubbers. However, the outlet duct has been a major problem area, particularly for units which have duct sections which handle both hot and wet gas. These sections are for the most part beyond the bypass junction on units which do not have reheat. Acidic conditions developed during scrubber operation become more severe on bypass as the temperature is raised and other corrosive species in the unscrubbed flue gas (chlorine and fluorine) are introduced.

Research efforts for these two components need to be directed to:

1. Compiling and maintaining general materials performance data
2. Characterizing environmental conditions where failures are occurring
3. Post-testing materials exposed to FGD environments to determine and/or verify failure mechanisms
4. Laboratory testing of commercial materials to verify proprietary data, and
5. Developing new or improved materials and designs based on the above information.

Prescrubbers, absorbers, reheaters, outlet ducts ahead of the outlet dampers, dampers, pumps, and piping and valves have a moderate history of materials problems but failures may not require complete boiler shutdown. Spray nozzles, mist eliminators, fans, inlet and bypass ducts, expansion joints, storage silos, ball mills, slakers, tanks, thickeners, agitators, rakes, vacuum filters, centrifuges, and pond linings have a relatively low history of materials problems and/or are amenable to rapid repair or replacement.

CRYSTALLIZATION OF GYPSUM
RP1031-2

OBJECTIVE

Forced oxidation with gypsum crystallization is being proposed as a means of solving or reducing the problems of handling sulfite sludge. Production of gypsum offers two alternatives to landfill sludge disposal. One alternative is to produce a gypsum of sufficient purity and consistency to be used in the manufacture of wallboard. The other is to produce gypsum of sufficient size to result in easy dewatering allowing "stacking" as another means of disposal.

In order to design FGD systems which will consistently produce a product of the desired properties, basic crystallization data are necessary. To obtain these data, a study of the nucleation rate and growth rate of gypsum has been completed. In addition, the effects of some operating conditions and additives on these properties were determined.

EXPERIMENTAL PROCEDURE:

Determination of nucleation and growth rates of gypsum were made in the "mini-nucleator" developed at the University of Arizona. The crystallizer in this apparatus is a one-liter, draft-tube-baffle, jacketed, glass vessel. Provisions are made to control temperature, liquor flow, and supersaturation. A particle counter by Particle Data, Inc. connected to a PDP-8 mini-computer is used to count particles and analyze data.

Supersaturation is normally developed by dissolving a desired compound at one temperature and crystallizing at a lower temperature for systems where solubility is temperature dependent. However, $CaSO_4$ has a low solubility and supersaturation was maintained by chemical reaction. Liquors were both simulated and actual limestone scrubbing liquors. Process variables studied were pH, agitation rate, and seed crystal concentration. The additives studies were sodium dodecylbenzenesulfonate, Calgon® CL246, adipic acid, and citric acid. These data were used in a computer simulation program to predict gypsum crystal size distribution from various crystallizer designs.

RESULTS AND CONCLUSIONS

Gypsum nucleates by secondary nucleation mechanisms of the collision breeding type when large (>150μm) parent crystals are retained in the crystal magma. High supersaturation and/or an absence of parent seed can result in bursts of excessive primary nuclei which degenerate particle size.

Although pH does not appear explicitly in the nucleation/growth rate kinetics expressions, the ratio of nucleation to growth shifts at low pH levels to produce smaller crystals. Regions of low pH (or sudden decrease in pH) in the scrubber system would be expected to reduce particle size.

Of the additives studied (sodium dodecylbenzenesulfonate, polyacrylate, adipic acid, citric acid), only citric acid had a beneficial effect on the size and shape of the crystals grown.

Computer simulations utilizing the nucleation/growth rate kinetics expressions developed in this study, together with assumed crystallizer configurations, indicate that particle size could be nearly doubled using a double-drawoff, classified removal crystalizer configuration in which mixed underflow and partially settled overflow streams are removed from the crystallization tank. Such operation could be achieved simply by installation of an internal settling baffle.

PLANS FOR FUTURE WORK

Since only one liquor composition was used in these studies, the effects of other ion concentrations (e.g., chloride and magnesium) in both lime and limestone system liquors will be studied. These data and the predicted size improvments are to be verified in a bench-scale crystallizer system.

CYCLIC REHEAT
RP1652-1

OBJECTIVE

A significant power plant operating cost savings is achievable if the cost of steam or oil used to reheat flue gas downstream of SO_2 wet scrubbing can be reduced or eliminated. One method of accomplishing this goal is by use of a cyclic reheat system which extracts heat from the flue gas entering the scrubber and uses that heat to reheat the stack gas. This study was performed to achieve a better understanding of cyclic reheat and to fill in information gaps regarding its application. The specific objectives are:

- To publicize the status of research work on cyclic reheat.
- To characterize the only existing U.S. full-scale cyclic reheat installation at Southwestern Public Service's (SPS) Harrington Station Unit 1 near Amarillo, Texas.
- To provide an economic comparison between cyclic reheat and conventional stack gas reheat schemes.

PROJECT DESCRIPTION:

The approach used to accomplish these objectives and develop the study information can be summarized as follows:

- Information on cyclic reheat research activities was obtained by literature search and by discussions with users, vendors, and research and engineering institutions regarding equipment types and systems used or considered for this application.
- Characterization of Harrington Station's cyclic reheat system was conducted by collecting historic design, cost, operating, and maintenance data; by performing gas sampling, component analyses, and temperature and pressure measurements on selected streams; and by analyzing these test data for system performance.
- Economic comparisons of cyclic reheat and three conventional stack gas reheat systems (in-line steam, hot-air injection, and oil-fired

reheat) were made on a comparable basis for both low and high sulfur coal cases. The comparisons are based on two 500 MW units operating with a reheat level of 50°F and with inlet flue gas (after dust collection) temperatures of 300°F and 270°F, respectively for the high and low sulfur cases. Materials of construction of exchanger tubes are chosen to take into account the sulfuric acid dewpoint and the temperature level of the reheat medium. Capital and operating costs are presented on a 30-year levelized basis. Cost sensitivity analyses were performed to determine the effect of certain design and energy value parameters.

RESULTS AND CONCLUSIONS

Results of the review of cyclic reheat research activities indicated that considerable effort has been and is being conducted on different approaches and equipment types. Most experience in the United States has been with gas/liquid/gas (Harrington) type systems, while in Japan it has been primarily with regenerative gas/gas (Ljungstrom) type cyclic reheat. Other approaches and equipment types in use or being studied include: (1) a heat pipe concept consisting of a closed tube containing a heat transfer medium which vaporizes during heat extraction and is condensed in reheating the scrubber outlet gas, (2) a borosilicate glass tube exchanger for gas/gas type cyclic reheat, and (3) a cast iron finned tube exchanger for low level heat recovery in a gas/liquid/gas system.

Characterization of cyclic reheat at Harrington Station which uses low-sulfur coal (0.3 to 0.5% sulfur) indicated superior operating experience with no serious corrosion or plugging problems despite carbon steel construction of the heat extractor and reheat exchangers. Results of the field test program indicated performance of the cyclic reheat and FGD systems are reasonably close to design. Sulfur trioxide (SO_3) content measured in the flue gas feed to the heat extractor was found to be considerably less than expected and indicates probable absorption and neutralization by the alkaline fly ash either in the flue gas or in the sampling system. Average finned area heat transfer coefficients for the heat extractor and reheat exchangers were found experimentally to be 6.3 and 10.3, respectively, as compared with values of

9 to 10 from accepted correlations when assuming the same flow conditions and a clean surface.

Results of screening-type capital estimates for both the Ljunstrom and heat pipe schemes indicated that these cyclic reheat approaches may have some cost advantage over the gas/liquid/gas system. However, a gas/liquid/gas system similar to that operating at SPS' Harrington Station and to that being installed at TVA's Paradise Steam Plant was chosen as a base case for comparison with conventional reheat methods because of the greater experience and availability of information.

Simplified EPRI Class 1 (±20%) capital and operating cost estimates were made for the gas/liquid/gas type cyclic reheat system and for the three conventional reheat systems, each with both high and low sulfur coal. Results are summarized in Table 2.

TABLE 2 REHEAT SYSTEM CAPITAL AND OPERATING COST SUMMARY

Basis: 2x500 MW coal-fired plant, Midwest location, 30-year plant life, pricing level-EOY 1979
HS - High-sulfur coal, 4.0% avg.
LS - Low-sulfur coal, 0.48% avg.
Capacity factor 70%

	Cyclic Reheat		In-Line Steam Reheat		Hot-Air In. Reheat		Oil-Fired Reheat	
	HS	LS	HS	LS	HS	LS	HS	LS
Process Capital, $/kW	17.7	22.8	6.2	6.8	3.1	3.2	2.2	2.4
Total Capital, $/kW	23.2	29.4	8.8	9.6	4.9	5.1	3.7	4.0
First Year O&M Cost, $/kW	1.9	2.2	4.1	4.4	9.5	10.6	5.0	5.5
Levelized Capital Charges, mills/kWH	0.68	0.86	0.26	0.28	0.14	0.15	0.11	0.12
Levelized O&M Cost, mills/kWH	0.60	0.69	1.30	1.42	3.04	3.40	1.97	2.17
Total 30-year Levelized Cost, mills/kWh	1.28	1.55	1.56	1.70	3.18	3.55	2.08	2.29

At the energy values of \$2.90 per 1000 pounds (\$6.39 per 1000 kg) for 70-250 psia (483-1724 kPa) steam, \$4.45 per 1000 pounds (\$9.80 per 1000 kg) for 250-600 psia (1724-4137 kPa) steam, \$4.40 per million Btu (\$4.17 per GJ) for oil, and 31 mills per kWh for electric power assumed in this study, cyclic reheat is estimated to have the lowest 30-year levelized total cost for stack gas reheat. Cyclic reheat has the highest capital requirement; direct combustion reheat has the lowest.

Study conclusions and recommendations include the following:

- Considerable research activities on cyclic reheat are being conducted with promising results. For low-sulfur coal application, operating experience has been good at the Harrington Station of Southwest Public Service. No serious corrosion or plugging problems are reported. For medium-sulfur coal application (1-2% sulfur) satisfactory operating experience is reported in Japan using the regenerative type of heat exchanger (Ljungstrom type). For high-sulfur coal application there is currently no operating experience; however, TVA's Paradise Steam Plant FGD system using cyclic reheat with high-sulfur coal is expected to start operation in 1982.

- The major advantage of cyclic reheat is energy savings. This is realized at the expense of higher capital cost. A simplified (EPRI Class I) estimate indicates that when medium-pressure steam costs \$2.30 per 1,000 pounds (\$5.10 per 1000 kg) or more, cyclic reheat has an economic advantage over conventional in-line steam reheat for high-sulfur coal. The breakeven point for low-sulfur coal application (based on an inlet flue gas temperature of 270°F (132°C) instead of 300°F (149°C)) is \$2.60 per 1,000 pounds (\$5.73 per 1000 kg).

- The capital cost of cyclic reheat is quite sensitive to the inlet flue gas temperature, which influences heat extractor size and materials of construction. Higher flue gas temperatures mean lower capital cost, but penalize power plant thermal efficiency. The comparative economics of a cyclic reheat system are also sensitive to the energy cost. Therefore, each plant should be independently evaluated.

- Cyclic reheat eliminates steam (or other fuels) consumption. However, this energy saving is partially offset by additional gas-side pressure drop across the exchanger surface. The exchanger size is influenced by the allowable pressure drop. High pressure drop improves heat transfer and reduces exchanger size, but consumes electrical energy in fan horsepower. Pressure drop and heat exchanger size must therefore be properly balanced to arrive at an economic optimum. In a gas/liquid/gas system, the design liquid temperatures must also be chosen to minimize the exchanger cost overall (extractors and reheaters).

- Cyclic reheat reduces scrubber inlet gas temperature. This has two effects on the main FGD system: (a) lowering the adiabatic saturation temperature of the gas, and (b) reducing the process water makeup requirement. Lowering the adiabatic saturation temperature may improve SO_2 removal efficiency, depending upon the particular FGD system. For the advanced concept of citrate absorption/steam stripping, a lower operating temperature means reduced steam consumption for SO_2 stripping (lbs steam/lbs SO_2). Reduced process water makeup may be beneficial in some arid areas; however, it also reduces the water available for mist eliminator wash. Both these factors are significant in FGD system design. Less water content in the scrubbed gas may enhance visibility by reducing the vapor plume.

- The rapidly escalating cost of energy has made cyclic reheat an increasingly attractive alternate to conventional reheat methods for FGD systems using wet scrubbing. However, before large-scale adoption of this reheat scheme takes place, several areas of uncertainty such as corrosion, plugging, and cleaning of the heat extractor should be investigated to minimize design errors and optimize equipment cost.

- Future studies should include in-depth studies of the Ljunstrom-type heat exchanger and of the heat pipe for cyclic reheat application. This would involve close monitoring of operating Ljunstrom-type systems in Japan.

ECONOMICS OF HIGH SO_2 REMOVAL DESIGNS

TPS 78-760, 78-767

OBJECTIVES

This project is a team effort by Radian Corporation and SRI International. The objective of Radian's work was to define representative cases and develop process designs and material balances that could be used to determine costs for each case. The process designs were developed using a process simulation computer program developed by the contractor. Cases were selected to span:

- Coal--eastern and western
- SO_2 removal--84%, 93% and 99%
- Alkali--magnesia, limestone and lime

The objective of SRI's work was to use the results of the Radian work to develop a cost estimate for each case and then analyze the results. The latest vendor cost information was used to prepare the economic estimates.

PROJECT DESCRIPTION

This project is composed of two separate technical planning studies that were undertaken to predict the effect of potential increasingly strict SO_2 emission limits on the economics of wet scrubbing. In the first study, Radian Corporation performed process designs and material balances as input to the second half of the study, an economic evaluation performed by SRI International.

RESULTS

Process Designs. The major variables that were investigated in these designs were the liquid-to-gas ratio (L/G) in the scrubber and the volume of the process slurry holding tank. The former affects the SO_2 removal efficiency and the latter affects the scaling potential in the scrubber.

Under the study assumptions, higher SO_2 removals required moderate increases in L/G and were found to be dependent on the magnesium and chloride levels in the slurry liquors. This information is useful in gaining an understanding of the magnitude of the process changes required for high SO_2 removals.

Cost Estimates. The study results are presented in Table 3. For low-sulfur coal systems, the design coal chosen meets the 1971 New Source Performance Standard (NSPS) for SO_2 without any further SO_2 removal. Increasing the design SO_2 removals to 93% and 99% results in a levelized cost of 8.5 and 8.9 mills/kWh, respectively. Magnesia scrubbing was about 7-8% more expensive than limestone scrubbing on a levelized basis for the low-sulfur western coal cases. For eastern higher-sulfur coal, increasing the removal requirements to 93% and 99% removal increases the levelized revenue requirement by 8% and 18%, respectively. Costs are significantly affected by chloride and magnesium levels in the coal. For high-sulfur coal, magnesia scrubbing is about 15% cheaper than limestone scrubbing on a levelized revenue basis.

The significance of the results of this study lies in the comparative numbers and not in their absolute magnitude. The increased costs are significant for higher SO_2 removals but they do not change by an order-of-magnitude as originally anticipated.

Probably the most significant unanticipated result of the study was the large effect that the Mg and Cl content of the scrubbing liquor has on system design and costs for lime and limestone systems. It is clear that this area should receive more attention in system design.

Finally, although the magnesia system appears less expensive than conventional lime and limestone systems for high-sulfur coals, it is still not well developed and its reliability remains uncertain.

Generalized cost estimates such as these are only an aid in planning either a research program or the selection of a flue gas desulfurization (FGD) process. It is not appropriate to generalize these comparisons or assume they represent manufacturers' current selling prices.

TABLE 3 ECONOMIC STUDY RESULTS

System	Percent SO_2 Removal	Type of Coal[1]	Levelized Revenue Requirement of FGD, Mills/kWh[2]	Total Capital Requirement, $/kW
Limestone	84	Eastern	13.0	165
Limestone[4]	93	Eastern	14.1	194
Limestone	99	Eastern	15.4	213
Limestone (High Cl)	93	Eastern	14.6	204
Limestone (Low Mg)	93	Eastern	13.8	189
Limestone (High Mg)	93	Eastern	12.9	178
Lime	93	Eastern	14.1	178
Limestone	93	Western	8.5	123
Limestone	99	Western	8.9	128
Magnesia[3]	93	Eastern	12.1	193
Magnesia	99	Eastern	13.1	207
Magnesia	93	Western	9.1	155
Magnesia	99	Western	9.6	163
Limestone	93	Eastern	14.4	181

1. Eastern coal, 4.0 sulfur; western coal, 0.48% sulfur; uncontrolled emissions would be 7.5 and 1.1 lb/million Btu, respectively. Eastern coal 0.1% Cl in base case, 0.3% in High Cl case.

2. Assuming an inflation rate of 6.0% per year and a fuel cost increase of 6.2% per year; 30-year levelized revenue requirement at levelized capacity factor of 0.7. Methodology standardized by EPRI.

3. Base cases.

4. Variation of base case design.

ECONOMICS OF FGD
RP1180-9

OBJECTIVES

The overall objective of this project was to prepare a general and consistent review of FGD technology economics. Specific objectives were to: (1) review reasons for variations between published FGD cost estimates, (2) recommend a consistent methodology for estimating FGD costs, and (3) prepare design and cost estimates for alternative FGD technologies using this methodology.

PROJECT DESCRIPTION:

An economic evaluation of flue gas desulfurization (FGD) processes was prepared by Bechtel National, Inc. The report presents a review of published FGD cost estimates, a discussion of the reasons for variations between published FGD costs, a recommendation of a methodology for improving the consistency of FGD cost estimates, and conceptual design and cost estimates for eight regenerable and nonregenerable FGD technologies, based on the recommended methodology.

FGD cost and performance estimates are presented for a new 2 x 500 megawatt unit plant located near Kenosha, Wisconsin and fired by either a 4-percent sulfur Illinois coal or a 0.48 percent sulfur Wyoming coal. Other major assumptions include 85 percent sulfur dioxide (SO_2) removal, four 33-1/3 percent scrubber modules, and redundancy in critical equipment. The evaluation was completed before promulgation of the final revised new-source performance standards for SO_2 in June 1979. Thus, the 70 to 90 percent SO_2 removal requirement was not used.

The FGD costs and other data presented in this report have also been used in a chapter on FGD economics in a report on sulfur oxides control technology being prepared by the National Research Council's Commission on Sociotechnical Systems.

EPRI intends to update and report FGD cost estimates on a regular basis, as technologies change.

RESULTS

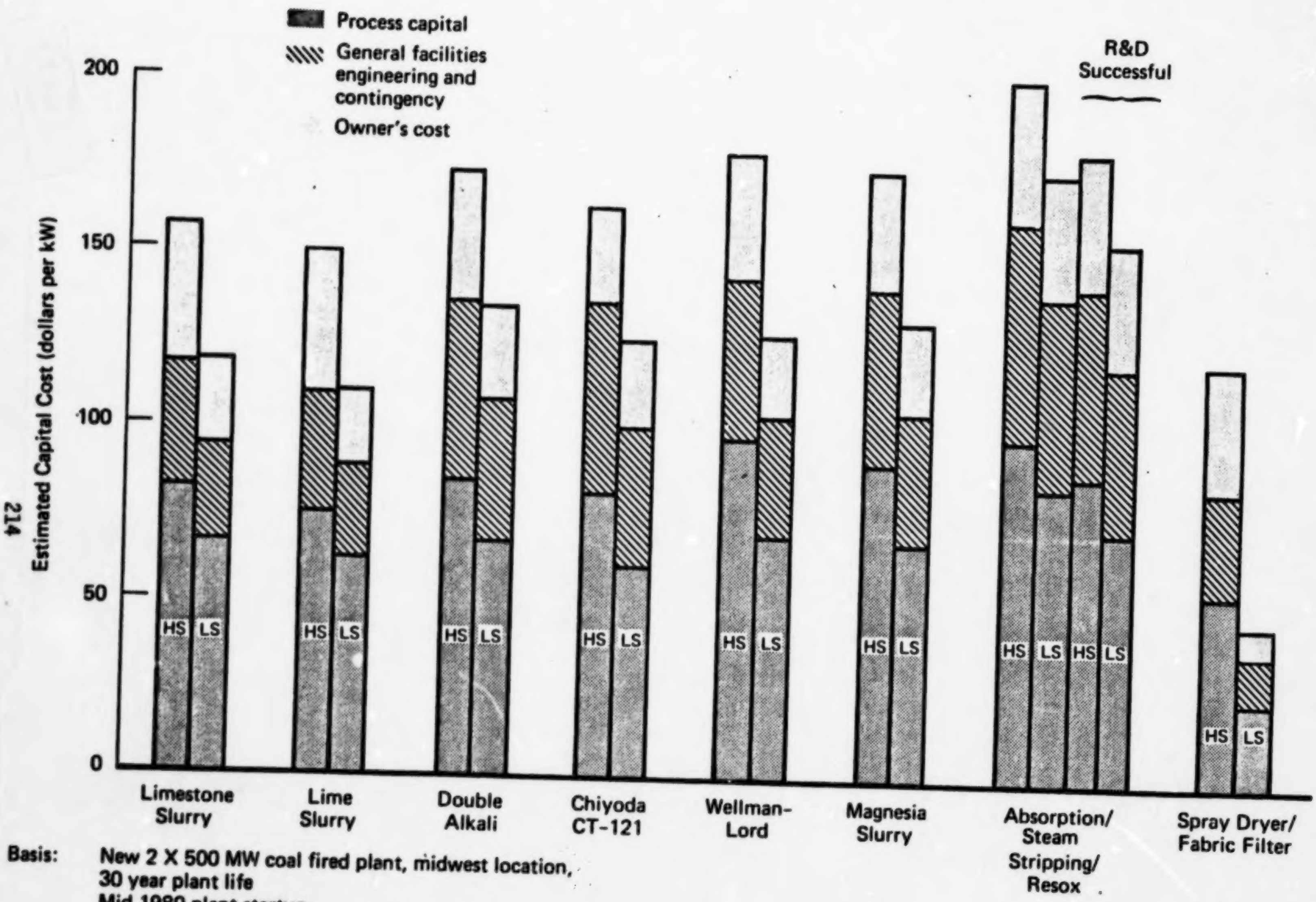
A review of nine published FGD cost estimates exhibited wide variations in both estimated and actual FGD costs. These variations often reach factors of three to five times the costs at the lower end of the cost range. The major causes of the cost variations are differences in SO_2 emission standard; scope of estimate; equipment redundancy; degree of design conservatism; purpose and level of detail of estimate; and design and economic assumptions, including coal type, plant location and capacity, and year of estimate.

The standard design and economic assumptions and methodology suggested in the report are expected to reduce the magnitudes of the differences between estimates. This methodology is already being used in other EPRI-sponsored FGD evaluations.

Conceptual designs and cost estimates are presented for the limestone slurry, lime slurry, double alkali, Chiyoda Thoroughbred 121, Wellman-Lord magnesia slurry, absorption/steam stripping/RESOX, and the lime slurry/spray drier/fabric filter processes.

For both low and high sulfur coal applications, the alkali-based non-regenerable processes exhibited the lowest capital and levelized revenue requirements and the lowest parasitic energy consumptions. The Chiyoda Thoroughbred 121 process appears particularly attractive. It exhibits low total capital and levelized revenue requirements and also produces a stackable gypsum byproduct in lower volumes than the sulfite sludge byproducts produced by the other limestone and lime slurry processes. The spray drier/fabric filter process using a lime slurry is also attractive, but has not yet been demonstrated for high sulfur coal applications. These costs are represented in the Figures 1 & 2.

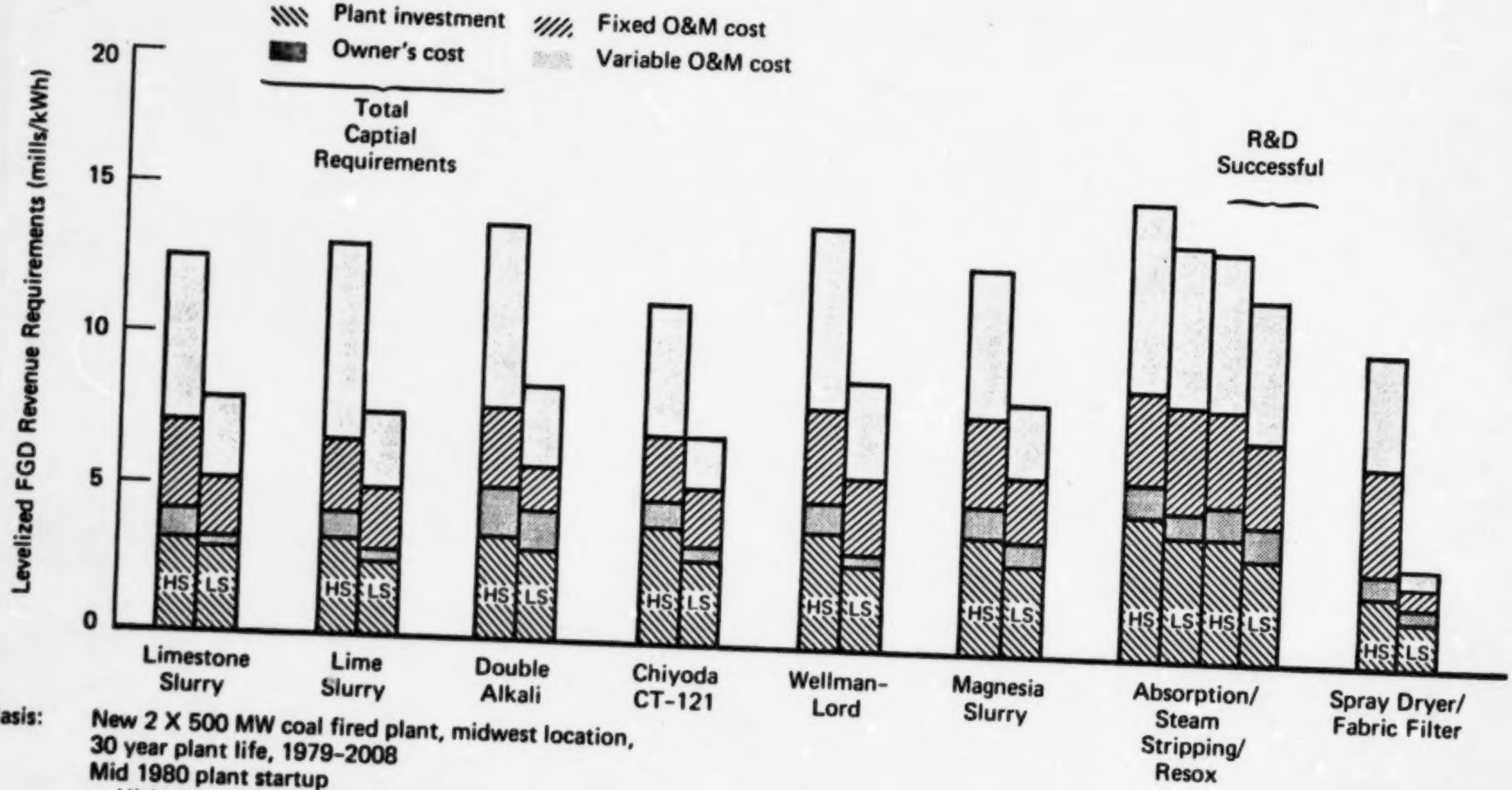
The economics of absorption/steam stripping and other regenerable processes are adversely affected by high energy consumption, principally for



Basis:
New 2 X 500 MW coal fired plant, midwest location,
30 year plant life
Mid 1980 plant startup
High sulfur coal - 4.0% sulfur (avg)
Low sulfur coal - 0.48% sulfur (avg)
Capacity factor 70%, 6132 hrs/yr
85% SO_2 removal

HS = High sulfur coal
LS = Low sulfur coal

Figure 1 Flue gas desulfurization–estimated capital cost.



Basis: New 2 X 500 MW coal fired plant, midwest location,
30 year plant life, 1979-2008
Mid 1980 plant startup
High sulfur coal - 4.0% sulfur (avg)
Low sulfur coal - 0.48% sulfur (avg)
Capacity factor 70%, 6132 hrs/yr
85% SO_2 removal

HS = High sulfur coal
LS = Low sulfur coal

Figure 2 30 Year levelized FGD revenue requirements, 1979-2008.

regeneration of the scrubbing reagent. Under RP1258, EPRI is evaluating improvements in the absorption/steam stripping/RESOX process that could reduce both energy consumption and equipment costs.

Generalized cost estimates such as those presented in this report should be used only as comparative estimates for research and development planning and FGD process screening. Since the estimates are based on a specific set of assumptions, it is not appropriate to generalize these estimates or assume they represent manufacturers' current, site-specific selling prices.

INTEGRATED EMISSION CONTROL PILOT PLANT
RP1646-1
RP1870-2

OBJECTIVE

The objective of EPRI's Integrated Emission Control (IEC) pilot plant research effort is to provide utilities with engineering guidelines for the specification of cost effective, reliable integrated emission control systems for coal fired plants.

PROJECT DESCRIPTION:

An integrated series of 2-1/2 MW pilot plant modules have been or are being constructed at EPRI's Arapahoe test facility in Denver, Colorado. The facility extracts flue gas from Public Service Company of Colorado coal fired unit. The catalytic NO_x control module and airheater are currently being tested. Additional modules to be tested include a spray dryer, a wet scrubbing system, a cooling tower, fabric filter and an electrostatic precipitator. The following elements or testing are planned:

- Complete characterization of each module and of integration effects.
- Implement a plant water chemistry program including integrating the water loop.
- Investigate impact of flue gas temperature.
- Determine effect of ammonia on air preheater, scrubber and fabric filters (baghouses).
- Test baghouse and ESPs as a final collection device.

RESULTS AND CONCLUSIONS:

The catalytic NO_x reactor has been operating since March 1980 with NO_x reduction performance close to original design. The test program is not far enough along to allow for detailed evaluation. Mr. Dan Giovanni, Program Manager of EPRI's Air Quality Control Program, can answer any general question on performance to date. The performance specifications for spray drying and wet scrubbing modules have been released for bids. A test sequence has been defined for several equipment configurations. All these activities amount to a multi-year R&D program that will represent the first attempt to integrate all of the best available control technologies into a single facility.

SULFUR PRODUCTION BY RESOX
RP784-2
RP1257-1

OBJECTIVE

The objective of EPRI's SO_2 reduction development efforts is to develop a regenerable FGD process that produces elemental sulfur without using a reducing gas such as methane (natural gas) or producer gas (CO,H_2). The RESOX process originally developed by Foster Wheeler Energy Corp. takes concentrated SO_2 produced by various FGD absorption systems and converts it to elemental sulfur by reaction with hot crushed coal.

PROJECT DESCRIPTION

Several related projects have contributed to the EPRI sponsored development of the RESOX technology. Early cost estimates developed for EPRI pointed to RESOX as a promising regeneration technique. The development effort is two fold with a U.S. 1 MW laboratory effort in Livingston, New Jersey, and a 42 MW demonstration effort in Lunen, Federal Republic of Germany. Initial German results were presented in a paper given in April 1979 at the ACS meeting in Honolulu.

RESULTS AND CONCLUSIONS:

Initial sulfur production was in July, 1978. Major equipment problems caused extended outages and little run time from August to March 1979. A total run time of approximately 900 hours was obtained, with the most productive runs in May and June of 1979. Low yields of 65-74% elemental sulfur based on a sulfur material balance caused EPRI to postpone further German efforts until problems were resolved in the lab. Lab runs were undertaken in October-November, 1979 attempting to reproduce German conditions and to find an improved method to correct the problem. Both goals were met in the lab program and the problem was diagnosed as overconversion of SO_2 to H_2S and COS. This was caused by an imbalance in gas flows, coal flow, and coal reactivity that led to high temperatures and, thus, low sulfur yields. The success of the improvement is causing EPRI to seek patent protection for the invention. Yields of 70% were recorded when reproducing Lunen conditions, and yields of 82.1, and 83.8% were recorded using the improved method of RESOX operation.

In order to increase the applicability of RESOX, additional lab work has been done to ascertain that bituminous or subbituminous coals as well as anthracite can be used as a reductant. Testing using these types of coals was performed with gases simulating Bergbau Forchung, Wellman-Lord, and magnesia off gas (Chemico-Basic). This testing was done without the improvement mentioned earlier, which leads us to believe that yields and sulfur purity can be increased. Even without the improvement, it still appears noncaking subbituminous and bituminous coals can be used in the RESOX process and that relatively dilute rich gas streams, such as magnesia off gas, can be processed. Coal types tested and results from this earlier testing are summarized in Table 4 & 5.

TABLE 4 RESOX TEST COALS

Mine/Seam	County/State	ASTM Ranking
Black Mesa/Yellow	Navajo, Arizona Bituminous	High Vol C
Seneca/Wadge	Routt, Colorado	Subbituminous A
Sophia Jacoba	Ruhr, Germany	Anthracite

TABLE 5 RESOX TEST RESULTS FOR VARIOUS COALS AND FRONT-END PROCESSES

	H_2O/SO_2 Mol Ratio	SO_2 in Feed (Mol %)	Inlet SO_2 Conversion (%)	Elemental Sulfur Yield (%)
Bergau-Forschung Process				
Sophia Jacoba coal	2.2	20.7	90.0	79.5
Black Mesa coal	2.2	12.0	92.1	85.2
Seneca coal	2.2	15.3	86.4	71.8
Wellman Lord Process				
Sophia Jacoba coal	2.5	24.4	91.3	80.0
Black Mesa coal	5.0	14.0	88.7	82.7
Seneca coal	6.0	11.8	84.5	75.0
Chemico-Basic Process				
Sophia Jacoba coal	5.0	8.3	91.6	79.7
Black Mesa coal	5.0	8.3	88.3	69.4
Seneca coal	4.0	9.0	82.6	68.1

WATER INTEGRATION SIMULATION FOR LIME AND LIMESTONE FGD SYSTEMS
TPS80-730

OBJECTIVE

Efficient utilization of water in power plants has become increasingly important particularly where water is scarce. For those plants which operate a wet scrubbing flue gas desulfurization (FGD) system, minimizing water usage requires careful study of overall water requirements with possible integration of water treatment and disposal in power plant and FGD systems. It may be possible, for example, to use some power plant waste streams in an FGD systems.

PROJECT DESCRIPTION

To determine the effects of various water streams on the operation of the FGD system, a computer model which calculates stream compositions for lime or limestone wet scrubbing has been used. This model will accept two different water compositions per material balance calculation. Approximately 40 different cases using raw water and waste streams such as cooling tower blowdown, and water treatment wastes in various combinations in the FGD systems have been done.

Four different raw water sources ranging in total dissolved solids (TDS) from 60 to 3400 ppm were chosen for study. Other variables are coal supply (one eastern and one western), FGD system (lime and limestone), prescrubber (with and without), and SO_2 removal efficiency.

These data were used in the Inorganic Process Simulator program of Radian Corp. to obtain reference material balances assuming that raw water was the only source of water for the FGD systems. Various plant streams (cooling tower blowdown, lime softening waste streams, etc.) were also calculated using these raw water sources in a computer program simulating cooling tower operations.

To determine if an FGD system could utilize any of the cooling tower waste streams as makeup water, combinations of raw water, cooling tower blowdown, and treatment wastes were used in the Inorganic Process Simulator was water sources. Material balances, scaling potential, operating conditions, scaling potential, operating conditions for the desired SO_2 removal, and stream compositions are determined by this computation program. Feasibility of using the cooling tower waste streams was judged by comparing the simulator waste stream data to those of the reference raw water data.

RESULTS AND CONCLUSIONS

Preliminary results of simulations using an eastern coal are shown in Table 6.

TABLE 6 Simulation of Desulfurization of Eastern Coal Flue Gas

SO_2 in Flue Gas	3000 ppm
SO_2 Removal Efficiency	90%
FGD Absorbent	Limestone

Makeup Water Source	TDS of Water, ppm	L/G Required		$CaSO_3$ Relative Saturation
		gal/kft^3	(l/m^3)	
Lake Sakajawea	3470	96	(12.8)	2.5
Santee River	66	129	(17.2)	1.4
Mississippi River	458	129	(17.2)	1.5
Cooling Tower Blowdown (Miss. River)	8460	86	(11.5)	2.8

Simulations using a western low-sulfur coal are given in Table 7.

TABLE 7 Simulation of Desulfurization of Western Coal Flue Gas

SO_2 in Flue Gas	400 ppm
SO_2 Removal Efficiency	70%
FGD Absorbent	Limestone

Makeup Water Source	TDS of Water, ppm	L/G Required		Absorber Effluent $CaSO_3$ Relative Saturation
		gal/kft^3	(l/m^3)	
Lake Sakajawea	3470	15	(2.0)	0.7
Mississippi River	458	62	(8.3)	0.1
Cooling Tower (Miss. River)	1530	53	(7.1)	0.5
Cooling Tower (Miss. River)	8480	25	(3.3)	1.8
Cooling Tower (Lake Sakajawea)	10,700	9	(1.2)	1.5

The preliminary results show that the quality of the water used can affect major variables such as L/G. Effects of water quality on lime slaking, lime and limestone availability and utilization, scaling, crystallization and mist elimination can also be indicated by these studies and wil be included in evaluations of the data as the work continues.

FUTURE PLANS

The simulator studies are to be continued using various combinations of water and waste water streams. Other coal, lime, and limestone compositions are to be combined in the system calculations. Build-up of impurities (such as chloride, sodium, and magnesium) will be calculated. Laboratory tests will then be completed to determine the effects, if any, on phase relationships, crystallization of calcium sulfite and sulfate, reagent utilization, and corrosion potential.

ENTRAINMENT IN WET STACKS
RP1653-1

OBJECTIVES

The history of wet stacks in the utility industry indicates two major problems relative to their operation: increased materials corrosion and mist generation. This project has been directed toward the problem of mist generation.

PROJECT DESCRIPTION

The work involves the collection and evaluation of historical data and laboratory pilot research on aerosol emission (reentrainment) from stack walls. The latter work involves experimental measurement of the critical velocity where water droplets are removed from the condensate film for the six different combinations of stack liner materials and construction roughness shown below.

STACK LINER MATERIALS TESTED FOR ENTRAINMENT FROM CONDENSATE FILM

1. Acid resistant brick (Custodis)
   Radical tolerance of construction (3.3×10^{-3}m or 0.13 in)
2. Acid resistant brick (Custodis)
   Radical tolerance of construction 0.0
3. SI units CXL - 2000 coating
   Colbran division of Pullman Power Products
4. Plastic coating No. 4005
   Vinyl Ester, Wisconsin Protective Coating Corp.
5. Inconel alloy welded
6. FRP (Fiberglass reinforced plastic) Alcore division of Custodis

Included in the study is an evaluation of choke design and operation on a wet stack. (The choke in a stack is the narrowing of the stack diameter at the top or exit to increase velocity and aid in plume dispersion.) This work includes experimental evaluation of two choke systems designed with water-film collectors. Separation and reentrainment prevention techniques for wet systems are also being evaluated using a mathematical model.

Based on the operating experience, mathematical modeling and experimental work, a set of guidelines for acceptable wet stack system designs will be formulated. The guidelines documents will include criteria for the selection of duct size and stack diameter, and a discussion for the trade-off between liner construction and critical reentrainment velocity, and the need for reentrainment prevention techniques or entrainment separation devices. The information is intended for use by A&E firms and utility owners to select or review wet stack system designs.

RESULTS AND CONCLUSIONS:

Tables 8 and 9 indicate some of the information that has been obtained in the survey of wet stack operating experience. Based on literature and laboratory measurements a properly operating mist eliminator carryover rate is 0.23-2.3 g/m^3 (0.1-1 gr/ft^3). Under upset conditions this can reach as high as 9.2 g/m^3 (4 gr/ft^3). Theoretical estimates of stack condensation range from 0.11-0.55 g/m^3 (0.05-0.24 gr/ft^3). If the measurements and estimates are accurate, the mist eliminator carryover is a significant portion of the condensed moisture in the stack and thus is a very important variable to be considered in the design of wet stack systems. The validity of these laboratory measurements needs to be confirmed by comprehensive field measurements.

TABLE 8 WET STACK DESCRIPTION

Plant Number	Entrainment Condition	No. Units	Particulate Removal	Absorber	I.D. Fans	Secondary Demisters
1	Entrainment is a big problem	2	Venturi	Mobile Bed	Dry	None
2	Moderate entrainment	1	ESP	Mobile Bed	Dry	None
3	Noticeable Entrainment during: - absorber overload - dirty demisters - plugged drains	2	Venturi	Venturi	Wet	Tried retrofitting 3 configurations with no success
4	No known entrainment	2	ESP	Venturi	Wet	Yes Chevron 4-pass
5	No known entrainment	2	ESP	Venturi	Wet	Yes Chevron 4-pass
6	No known entrainment	2	ESP	Mobile Bed	Dry	None
7	Slight noticeable entrainment during humid weather	2	ESP	Mobile Bed	Dry	None

TABLE 9 STACK BREECHING DUCT DATA

Stack Data							Breeching Duct Data - Entrance to Stack			
Plant	Height	Diameter (base-top)	Gas Velocity (max)	Gas Temperature (average)	Liner Material	Liner Insulation	No. Flues	Height	Width	Distance from dust to stack exit
1	50m (165 ft)	4.9m (16 ft.)	7.6m/s (25 ft/s)	43.3°C (110°F)	Carbon Steel w/ Precrete	None	1	Scrubber & demister share stack flow enters at bottom of stack.		From demister 23m (75 ft)
2	183m (600 ft)	5.9m (19-1/2 ft)	27.4 m/s (90 ft/s)	48.9°C (120°F)	Mild steel w/ Ceilcoat	2-3 inches fiberglass	1			
3	290m (950 ft)	5.8m (19 ft)	27.4 m/s (90 ft/s*)	48.9°C (120°F)	Carbon Steel w/ Heil Rigi-flake	None	4	7.6m (25 ft)	3.7m (12 ft)	247m (810 ft)
4	119m (390 ft)	8.8-7.9m (29-26ft)	7.3m/s (24 ft/s)	48.9°C (120°F)	Acid proof Brick & Mortar	None	1	12.2m (40 ft)	3.7m (12 ft)	88m (290 ft)
5	104m (340 ft)	8.8-7.9m (29-26 ft)	11m/s (36 ft/s)	48.9°C (120°F)	Acid proof Brick & Mortar	None	1	12.2m (40 ft)	3.7m (12 ft)	72m (2325 ft)
6	244m (800 ft)	13.4-7.9m (44-26 ft)	24.4 m/s (80 ft/s)	51.7°C (125°F)	Acid Proof Brick & Mortar	None	1	9.1m (30 ft)	4.6m (15 ft)	213m (700 ft)
7	137m (450 ft)	3.4m (11 ft)	30 m/s (95 ft/s)	54.4°C (130°F)	Acid Proof Brick & Mortar	None	1	5.8m (19 ft)	2.3m (7-1/2 ft)	101m (330 ft)

* Secondary source gives 60 ft/sec velocity.

FUTURE PROGRAM EMPHASIS

Future R&D emphasis will be on the following:

- Demonstrations
  - Chiyoda 121
  - RESOX
  - Aqueous carbonate process
- Pilot Plant
  - Absorption/steam stripping improvements
  - Spray dryer testing
  - Integrated emission control
- Field Testing
  - Continuous emission monitor testing
  - Materials testing
  - Spray dryer chacterization
- Evaluations
  - Reliability improvements
  - Cyclic reheat feasibility
- Laboratory Testing
  - Corrosion inhibitors
  - Limestone dissolution
  - Crystallization
  - Additives

- Tech Transfer
  - -- Revised Lime FGD Systems Data Book
  - -- Issue limestone data book
  - -- Continuous emission monitor guidelines
  - -- Workshops and seminars

CONCLUSION

EPRI research and development has attempted to address problems in FGD which have led to the high cost, low reliability and inefficient resource use in current systems. EPRI's efforts are aimed at near term solutions to problems in system chemistry, corrosion, cost, energy use, by-product character, and system design. The results reported in this paper are documented more fully in individual reports that are either in print or in the process of publication. EPRI welcomes and encourages comments, criticisms or inquiry regarding its FGD programs and asks that such calls be directed to Stu Dalton, Program Manager, Desulfurization Processes Program. (415) 855-2467.

Session 4: UTILITY APPLICATIONS

H. William Elder, Chairman
Tennessee Valley Authority
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TEST RESULTS ON ADIPIC ACID-ENHANCED LIMESTONE SCRUBBING AT THE EPA SHAWNEE TEST FACILITY

-THIRD REPORT-

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ABSTRACT

Adipic acid has been demonstrated as a powerful scrubbing additive for enhancing SO_2 removal in lime and limestone wet scrubbing tests both at the EPA/IERL pilot plant at Research Triangle Park, North Carolina, and at the EPA-sponsored Shawnee Test Facility near Paducah, Kentucky. Improved limestone utilization and operating reliability have also been demonstrated.

Earlier test results using adipic acid, from July 1978 through October 1979, were reported at the Fifth Symposium on Flue Gas Desulfurization in Las Vegas, Nevada, March 5-8, 1979, and at EPA's Fifth Industry Briefing in Raleigh, North Carolina, December 5, 1979. This is the third report on the recent adipic acid test results at the Shawnee Test Facility from October 1979 through May 1980.

The recent tests with adipic acid were conducted only on the venturi/spray tower system. All tests were made with limestone slurry. These included: (1) partial factorial tests to characterize the effects of pH, adipic acid concentration, and other operating parameters on SO_2 removal; (2) single-loop (one-tank) tests without forced oxidation at low pH and high (4000 ppm) adipic acid concentration; (3) tests with a venturi only to determine the limits of SO_2 removal; (4) single-loop forced oxidation tests, with both one tank and two tanks; and (5) bleed stream oxidation tests at low pH and high (4000 ppm) adipic acid concentration.

Major efforts during the recent test period were directed toward investigation of the effect of pH on the degradation of adipic acid. It was found that the adipic acid degradation is minimized when the scrubber is operated at low (below 5.0-5.1) inlet pH. Forced oxidation and poor limestone utilization tend to increase the degradation.

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NOTE

Although it is the policy of the EPA to use the metric system for quantitative descriptions, the British system is used in this report. Readers who are more accustomed to metric units are referred to the conversion table in the Appendix.

TEST RESULTS ON ADIPIC ACID-ENHANCED LIMESTONE SCRUBBING AT THE EPA SHAWNEE TEST FACILITY

- THIRD REPORT -

Section 1

INTRODUCTION

Since October 1977 one of the primary objectives of the Environmental Protection Agency (EPA) alkali wet scrubbing test program has been to enhance SO_2 removal and improve the reliability and economics of the lime and limestone wet scrubbing systems by use of adipic acid as a chemical additive.

Testing of adipic acid-enhanced limestone scrubbing began in October 1977 at the EPA 0.1 MW pilot plant at Industrial Environmental Research Laboratory, Research Triangle Park (IERL-RTP), North Carolina (Reference 1). As a logical progression, larger scale testing was conducted beginning in July 1978 at EPA's 10 MW prototype Shawnee Test Facility located at the Tennessee Valley Authority (TVA) Shawnee Steam Plant near Paducah, Kentucky. Test results from the Shawnee Test Facility from July 1978 through October 1979 were presented in two previous reports (References 2 and 3). As part of EPA's continuing program of technology transfer, to further demonstrate the effectiveness of adipic acid, and to encourage its use, the EPA contracted with Radian Corporation in the spring of 1980 to carry out a full-scale demonstration program of adipic acid-enhanced limestone scrubbing. The program is being conducted at the Springfield City Utilities' Southwest Station near Springfield, Missouri. Testing in the full-scale units began in the late summer of 1980.

This report is the third presenting the test results with adipic acid from the Shawnee Test Facility. The report covers the period from October 1979 through May 1980. During this period, adipic acid testing was conducted only on the venturi/spray tower system (Train 100). All tests were conducted with limestone slurry and with flue gas having high fly ash loading (3 to 6 grains/scf dry).

During the report period, Train 200 (TCA) was operated by EPRI/UOP/TVA on a DOWA basic aluminum sulfate process, and Train 300 was operated by EPRI/TVA on a cocurrent, high-velocity scrubber configuration.

THEORY AND ADVANTAGES OF ADIPIC ACID-ENHANCED SCRUBBING

Adipic acid is a dicarboxylic organic acid [$HOOC(CH_2)_4COOH$] in powder form, which is commercially available and used primarily as a raw material in the nylon manufacturing industry. Initial tests with adipic acid at the

IERL-RTP pilot plant were undertaken as a result of theoretical analyses by Rochelle (Reference 4). Adipic acid effectively buffers the pH in limestone/lime SO_2 absorbers and improves the SO_2 removal efficiency. The buffering action limits the drop in pH at the gas/liquid interface during SO_2 absorption, and the resultant higher concentration of SO_2 at the interface accelerates the liquid-phase mass transfer. The capacity of the bulk liquor for reaction with SO_2 is also increased by the presence of calcium adipate in solution. Thus, the SO_2 absorption becomes less dependent on the dissolution rate of limestone or $CaSO_3$ in the absorber to provide the necessary alkalinity. In the case of limestone scrubbing, it follows that a given SO_2 removal efficiency can be achieved at a lower limestone stoichiometry.

Further analysis by Rochelle (Reference 5) suggested that the use of additives would be most attractive economically when used in scrubbers employing forced oxidation. If no decomposition or volatilization of the additive occurs, the makeup requirements of the additive would be minimized by the more tightly closed liquor loop achievable due to the better dewatering properties of the oxidized sludge.

Several advantages of adipic acid over other additives, such as MgO, have been cited previously (References 1, 2, and 3). Further, the optimum concentration of adipic acid for effective improvement in SO_2 removal is only 700 to 1500 ppm at a scrubber inlet pH above about 5.2. The preliminary economic evaluations (Reference 2) have shown that adipic acid can reduce both the capital investment and the operating cost of limestone systems while simultaneously increasing the performance, even under those conditions in which the actual addition rate is 3 to 5 times the theoretical requirement due to the degradation of the acid.

This report shows that the degradation of adipic acid can be minimized when the scrubber inlet pH is lowered to below about 5.0. Although higher adipic acid concentration is needed at the lower pH to achieve the same degree of SO_2 removal efficiency, overall adipic acid consumption is reduced compared to the higher pH operation. For this reason, and with the further improvement in limestone utilization at low pH, the low pH operation should be more economically attractive. Section 11 presents an update of the economic evaluations given in Reference 2.

TEST FACILITY AND PROGRAM

Readers who are unfamiliar with the Shawnee Test Facility and the earlier adipic acid test programs are referred to References 2 and 3. A summary of the earlier work is given in Section 2. This report covers the adipic acid test results from October 1979 through May 1980 on the venturi/spray tower system. The following adipic acid tests were conducted during this period:

- Partial factorial tests to characterize the venturi/spray tower performance using a single tank without forced oxidation
- Investigation of the effect of pH on adipic acid degradation with and without forced oxidation

- SO_2 removal capability of the venturi scrubber alone
- Forced oxidation within the scrubber loop using a single tank
- Forced oxidation within the scrubber loop using two tanks in series
- Forced oxidation of the bleed stream

All tests were conducted using limestone slurry and flue gas containing 3 to 6 grains/dry scf of fly ash. Sections 3 to 8 discuss and summarize these tests.

Section 9 describes scrubber system behavior during limestone blinding and the conditions leading to it. Recommended solutions for eliminating or avoiding limestone blinding are also given. Section 10 gives updated data on the dewatering properties of adipic acid-enhanced limestone slurry.

Section 2

SUMMARY OF PREVIOUS WORK

Based on the earlier test results through October 1979 (References 1, 2 and 3), both at the IERL-RTP pilot plant and at the Shawnee Test Facility, the characteristics of adipic acid as a lime/limestone scrubber additive can be summarized as follows:

BENEFICIAL ASPECTS

- Adipic acid significantly enhances SO_2 removal. At a scrubber inlet pH above about 5.2, at which most of the adipic acid is in ionized form, the optimum concentration range is only 700 to 1500 ppm.
- At the minimum effective pH of 5.2, the corresponding limestone utilization is normally about 80 percent or higher; thus the quantity of waste solids generated is reduced. High limestone utilization also contributes to reliable scrubber operation.
- With proper pH control and sufficient adipic acid concentration (sufficient buffer capacity), steady outlet SO_2 concentrations can be maintained even with wide fluctuations of inlet SO_2 concentrations.
- Adipic acid-enhanced limestone scrubbing has lower projected capital and operating costs than unenhanced limestone or limestone/MgO scrubbing (Reference 2). This is primarily due to the reduced limestone consumption, the associated grinding cost, and the reduced quantity of waste sludge generated with adipic acid-enhanced scrubbing.
- Since limestone dissolution is not a rate-controlling step in SO_2 absorption for an adipic acid-enhanced limestone system, adipic acid should promote use of less expensive and less energy-intensive limestone than lime.
- The effectiveness of adipic acid is not affected by forced oxidation and it can be used with both lime and limestone in systems with or without forced oxidation.
- The effectiveness of adipic acid is not adversely affected by chlorides as is the limestone/MgO process. Thus it is especially attractive for very tightly closed liquor-loop operation.
- When used with lime, both good SO_2 removal and sulfite oxidation can be achieved in a single-loop scrubbing system using within-scrubber-loop forced oxidation.

NEGATIVE ASPECTS

- Adipic acid decomposition, and the indications of its being adsorbed on solids or occluded in solids (Reference 6), require adding up to

5 times that amount theoretically required (Reference 2). However, the consumption over the ranges anticipated has negligible economic impact.

- Some decomposition products, such as valeric acid, have an unpleasant odor. However, this has not been a problem in testing to date.

OTHER CONSIDERATIONS

Toxicity. No further work in this area has been conducted by the EPA since the last report (Reference 3). Preliminary results from Level 1 chemical and bioassay analyses showed no measurable difference in toxicity or mutagenicity of samples with and without adipic acid addition. These samples were taken in February 1979 from a limestone/adipic acid forced-oxidation run and a base case limestone run without forced oxidation. It should be noted that adipic acid is a food additive.

Limestone Blinding and Calcium Sulfite Scaling. Adipic acid buffers the pH drop across the scrubber, and therefore increases the potential of calcium sulfite scaling at the bottom part of the scrubber. At a constant liquid-to-gas ratio, addition of adipic acid increases the SO_2 make-per-pass and similarly increases the sulfite scaling tendency at the bottom of the scrubber. In the case of limestone scrubbing, blinding of limestone by calcium sulfite could occur, resulting in low pH and poor limestone utilization. This would be particularly true with forced oxidation in the scrubber loop (or in a system with a high level of natural oxidation); such conditions deplete calcium sulfite solid seeds. Operating and design considerations for avoiding limestone blinding are presented in Section 9.

Section 3

FACTORIAL TESTS ON THE VENTURI/SPRAY TOWER SYSTEM WITH LIMESTONE/ADIPIC ACID SLURRY

Fifty limestone/adipic acid partial factorial tests, Runs VAA201 through VAA250, were conducted on the venturi/spray tower system. All tests were made without forced oxidation and with a common effluent hold tank as shown in Figure 3-1.

The tests examined the effect of spray tower liquid-to-gas ratio, scrubber inlet liquor pH, and adipic acid concentration on SO_2 removal. Table 3-1 summarizes the test results. The operating conditions held constant during these tests were:

Fly ash loading: High (3-6 grains/dry scf)
Flue gas rate: 35,000 acfm @ 300°F (except Run VAA 207 @ 20,000 acfm)
Hold tank level: 8 ft 6 in. (9.1 - 38 minutes residence time)
Slurry solids concentration: 15 percent
Venturi pressure drop: 9 inches H_2O for runs with 600 gpm, plug wide open for runs with 125 gpm
Spray header configuration (top header is No. 4):
  For 400 gpm ---- Header 4
  For 800 gpm ---- Headers 3 and 4
  For 1200 gpm ---- Headers 2,3, and 4
  For 1600 gpm ---- All four headers
Solids dewatering equipment: Clarifier and centrifuge

OVERALL SO_2 REMOVAL BY VENTURI AND SPRAY TOWER

Equation 3-1 for predicting SO_2 removal has been fitted to the 10 venturi/spray tower runs (Runs VAA201 through VAA206 and VAA234 through VAA237) for which the slurry flow rate to the venturi was held at 600 gpm and the venturi pressure drop was 9 inches H_2O.

$$\text{Fraction } SO_2 \text{ Removal} = 1 - \exp\left[-0.0019\,(L/G)^{0.55}\,\exp(0.8\text{pH} + 8\text{x}10^{-4}\,A)\right] \quad (3\text{-}1)$$

where:

L/G = spray tower liquid-to-gas ratio, gal/mcf (saturated)
pH = scrubber inlet liquor pH
A = adipic acid concentration in scrubber liquor, ppm

The ranges of operating variables covered by the 10 correlated runs are:

L/G = 15-57 gal/mcf
pH = 5.2-5.8 (limestone stoichiometry controlled at 1.2)
A = 600-1400 ppm
Gas flow rate = 35,000 acfm at 300°F
Inlet SO_2 concentration = 1500-2200 ppm

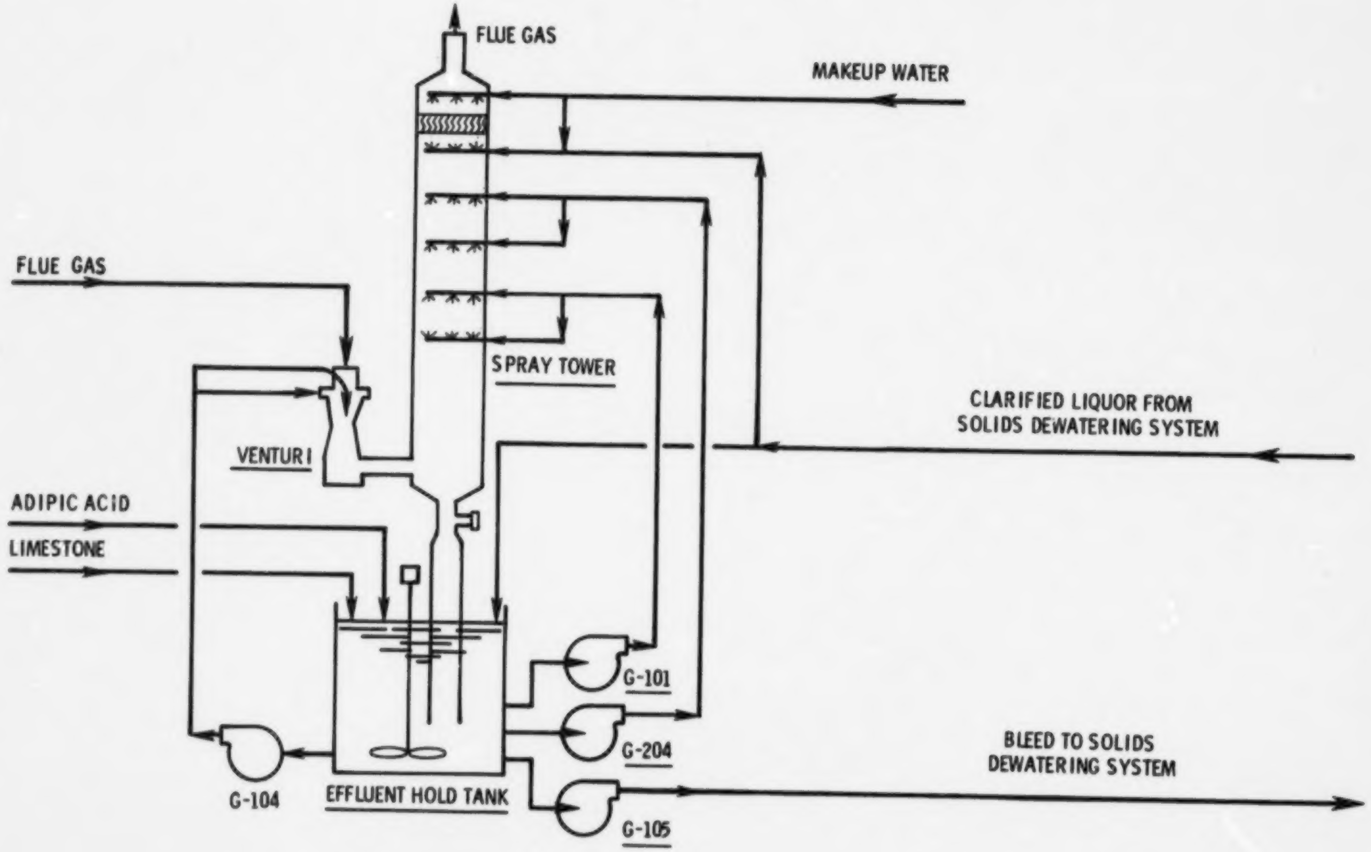

FLUE GAS
MAKEUP WATER
FLUE GAS
SPRAY TOWER
CLARIFIED LIQUOR FROM
SOLIDS DEWATERING SYSTEM
VENTURI
ADIPIC ACID
LIMESTONE
G-101
G-204
BLEED TO SOLIDS
DEWATERING SYSTEM
G-104
EFFLUENT HOLD TANK
G-105

Figure 3-1. Flow Diagram for the Venturi/Spray Tower System With One Tank and Without Forced Oxidation

Table 3-1

RESULTS OF LIMESTONE/ADIPIC ACID FACTORIAL TESTING ON THE VENTURI/SPRAY TOWER USING ONE TANK WITHOUT FORCED OXIDATION

Run No.	Liquor Rate (gpm)		ST L/G (gal/mcf)	Pressure Drop (inch H_2O)		Inlet Liquor pH	Outlet Liquor pH	Ave AD* Conc. (ppm)	Ave. Cl^- Conc. (ppm)	Inlet SO_2 Concentration (ppm)		Percent SO_2 Removal		Venturi SO_2 Removal	Replicate
	Venturi	ST		Venturi	Total					Range	Ave.	Range	Ave.		
VAA201	600	1200	43	8.9	13.5	5.70	5.5	710	2410	1880-2120	2005	90-95	93.1	-	A
VAA202	600	800	29	8.7	14.4	5.45	5.2	640	2880	1760-2000	1850	79-84	81.6	45.5	B
VAA203	600	400	15	8.8	13.6	5.85	5.6	590	1240	1680-1840	1760	73-76	74.4	-	
VAA204	600	1600	57	8.9	14.1	5.75	5.4	690	1850	1480-1520	1505	95-96	95.8	-	
VAA205	600	800	29	8.9	15.0	5.80	5.6	650	1910	1560-1600	1570	86-91	88.9	-	B
VAA206	600	1200	43	8.7	14.6	5.65	5.5	680	2710	1760-1880	1850	84-90	86.4	-	A
VAA207	125	1200	75(1)	0.9	3.1	5.75	5.6	570	1740	2000-2080	2045	92-93	92.4	-	
VAA208	125	800	29	2.3	7.7	5.85	5.6	610	1810	1840-1980	1910	72-80	76.4	-	C
VAA209	125	1200	43	3.1	7.9	5.70	5.4	550	1720	2120-2160	2150	77-78	77.9	-	D
VAA210	125	400	15	2.4	7.0	5.80	5.5	620	1880	1600-1760	1715	53-58	54.7	20.6	
VAA211	125	1600	57	3.7	8.8	5.85	5.6	730	1860	1740-2100	1920	87-88	87.3	-	
VAA212	125	1200	43	2.6	7.9	5.60	5.3	650	2080	1760-2000	1835	79-83	80.7	-	D
VAA213	125	800	29	2.5	7.7	5.65	5.5	660	1600	1840-2000	1965	69-74	72.1	17.6	C
VAA214	125	1200	43	3.1	8.6	5.35	5.0	660	2170	1800-2200	2010	68-76	71.4	-	
VAA215	125	1200	43	2.7	9.0	5.05	-	610	2120	1760-2200	2115	57-65	60.5	-	E
VAA216	125	1200	43	2.7	8.1	4.65	5.2	660	1810	1800-2000	1900	44-50	46.0	-	
VAA217	125	1600	57	2.7	7.7	5.35	5.6	590	2170	1720-2000	1855	71-83	79.2	-	
VAA218	125	1600	57	2.3	7.6	5.00	4.7	660	2280	2080-2280	2165	54-63	58.3	-	
VAA219	125	1200	43	2.6	8.0	5.00	-	710	2240	1760-2160	1940	56-62	58.0	-	E
VAA220	125	1600	57	2.7	7.4	4.55	-	840	2310	2040-2080	2075	36-42	40.3	-	
VAA221	125	800	29	1.9	8.6	5.50	-	1170	2640	1880-2000	1910	72-76	75.2	-	F
VAA222	125	400	15	3.3	7.9	5.50	-	1290	3250	1960-2000	1985	55-61	57.9	21.9	G
VAA223	125	1200	43	3.0	7.8	5.40	5.4	1370	3080	1800-2080	1945	83-88	86.5	-	
VAA224	125	1600	57	3.4	8.2	5.50	-	1190	3120	1920-2060	1900	91-93	92.6	-	
VAA225	125	800	29	3.1	7.9	5.45	5.3	1190	2930	1720-1920	1840	79-83	81.2	-	F
VAA226	125	400	15	4.0	9.0	5.60	-	1280	4050	1740-2040	1880	63-66	64.6	-	G
VAA227	125	1200	43	3.1	8.0	5.40	-	1270	2900	1400-1680	1595	87-91	89.8	-	
VAA228	125	1200	43	3.6	8.9	4.90	-	1450	2990	1880-2040	1930	66-70	69.2	-	H
VAA229	125	1200	43	3.7	8.9	4.70	-	1380	3370	1640-1780	1735	50-55	51.9	-	
VAA230	125	800	29	4.6	9.7	5.45	-	1350	3470	1880-2000	1970	69-80	77.7	-	
VAA231	125	800	29	4.1	9.6	5.05	-	1340	3240	1740-2200	1900	59-72	64.9	21.6	
VAA232	125	1200	43	4.2	8.9	5.00	-	1410	4410	2260-2560	2400	69-80	74.7	-	H
VAA233	125	800	29	3.9	8.9	4.65	-	1400	3700	1720-1880	1780	45-50	48.4	-	
VAA234	600	800	29	9.0	14.0	5.20	-	1300	4170	2000-2200	2155	85-90	87.4	51.9	I
VAA235	600	400	15	9.2	15.2	5.45	-	1350	4030	1520-1800	1640	84-89	85.8	58.0	
VAA236	600	1200	43	8.9	12.2	5.35	-	1410	4180	1960-2240	2145	95-97	95.9	-	
VAA237	600	800	29	9.0	11.3	5.40	-	1290	4260	1860-2040	1915	93-94	93.6	-	I
VAA238	125	800	29	3.1	8.1	5.30	-	2310	4580	2480-2720	2625	84-90	88.3	39.4	J
VAA239	125	1200	43	3.2	8.1	5.65	-	1980	4110	2340-2520	2415	96-97	96.5	42.4	K
VAA240	125	400	15	3.7	8.5	5.25	4.8	1900	4000	2800-3000	2935	68-71	69.3	46.0	L
VAA241	125	1600	57	3.7	9.3	5.35	4.9	1810	3760	1840-2240	2025	96-97	96.6	-	
VAA242	125	800	29	3.1	7.6	5.65	-	2070	3930	2240-2400	2340	93-95	94.3	-	J
VAA243	125	1200	43	2.9	7.7	5.60	-	2040	3850	2360-2480	2430	96-97	96.8	-	K
VAA244	125	400	15	3.8	8.6	5.45	4.8	2300	5030	2200-2600	2480	68-72	69.5	-	L
VAA245	125	800	29	3.0	7.5	5.40	-	2260	3690	2360-2640	2440	91-93	92.1	-	
VAA246	125	800	28	3.0	7.9	5.05	-	2370	3850	1800-2160	1900	87-93	89.9	-	
VAA247	125	800	29	3.8	9.0	4.60	4.5	2170	4010	2660-2880	2810	60-66	63.5	-	
VAA248	125	1200	43	2.9	7.7	5.50	-	2000	3620	1840-2160	1980	96-98	96.8	-	
VAA249	125	1200	43	3.0	7.7	5.00	-	2270	4640	2180-2360	2245	90-93	91.7	-	
VAA250	125	1200	43	3.8	9.2	4.55	4.2	2210	4440	2300-2560	2470	69-73	71.3	-	

(1) Flue gas rate = 20,000 acfm @ 300°F for Run VAA207.
All other runs at 35,000 acfm @ 300°F.

Venturi liquid-to-gas ratio = 21 gal/mcf
Venturi pressure drop = 9 inches H_2O

Equation 3-1 explains 90 percent of the variation in the data for SO_2 removal with a standard error of estimate of 2.7 percent SO_2 removal (see Figure 3-2).

SO_2 REMOVAL BY SPRAY TOWER ONLY

Equation 3-2 for prediction of SO_2 removal has been fitted to the 40 spray tower runs (minimum effect of venturi - 125 gpm for flue gas humidification):

$$\text{Fraction } SO_2 \text{ Removal} = 1 - \exp\left[-2.2\times10^{-4}\,(L/G)^{0.75}\,\exp\left(pH + 6.2\times10^{-4}\,A\right)\right] \quad (3\text{-}2)$$

where L/G, pH, and A have the same definitions as for Equation 3-1.

The ranges of variables covered by the 40 correlated runs are:

L/G = 15-75 gal/mcf
pH = 4.6-5.9
A = 600-2400 ppm
Gas flow rate = 35,000 acfm at 300°F (one test at 20,000 acfm)
Inlet SO_2 concentration = 1600-2900 ppm
Venturi slurry flow rate = 125 gpm
Venturi pressure drop = 2-4 inches H_2O (wide open plug)

Equation 3-2 explains 93 percent of the variation in the data for SO_2 removal with a standard error of estimate of 4.3 percent SO_2 removal (see Figure 3-3). It is important to note that the SO_2 removal predicted by Equation 3-2 includes the effect of the venturi operating at the minimum conditions defined above. The magnitude of this effect is discussed later.

Figures 3-4 through 3-6 illustrate the effects of spray tower liquid-to-gas ratio and inlet liquor pH on SO_2 removal at adipic acid concentrations of 600, 1300, and 2000 ppm, respectively. The lines on the figures represent the predictions of Equation 3-2 with actual data points also shown.

Note that the SO_2 removals for a pH of 4.6 and 2000 ppm adipic acid in Figure 3-6 are similar both to those in Figure 3-5 for a pH of 5.0 and 1300 ppm adipic acid, and to those in Figure 3-4 for a pH of 5.4 and 600 ppm adipic acid. These values are more clearly seen in the following:

Scrubber Inlet pH	Adipic Acid, ppm	Percent SO_2 Removal at Spray Tower L/G of		
		30 gal/mcf	50 gal/mcf	70 gal/mcf
5.4	600	60	74	82
5.0	1300	61	75	83
4.6	2000	62	76	84

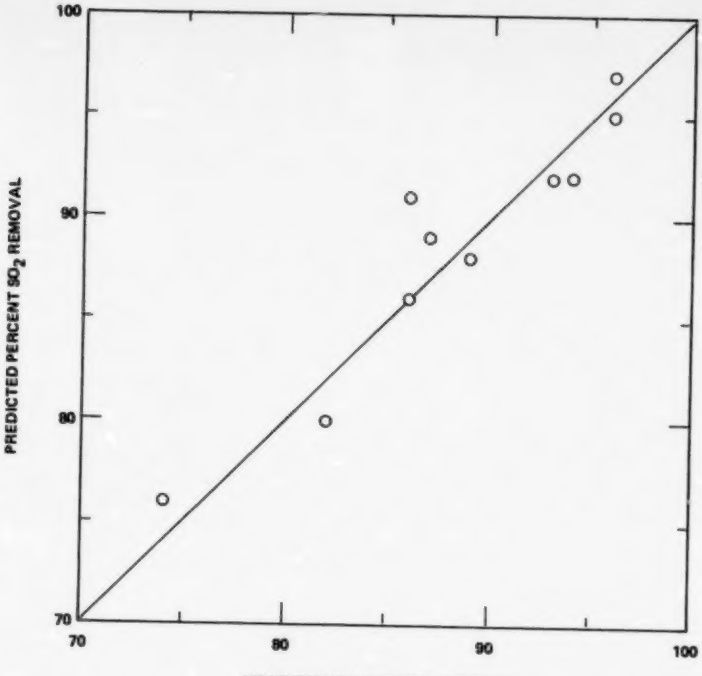


Figure 3-2. Measured vs. Predicted (Eq. 3-1) SO_2 Removal by the Venturi/Spray Tower

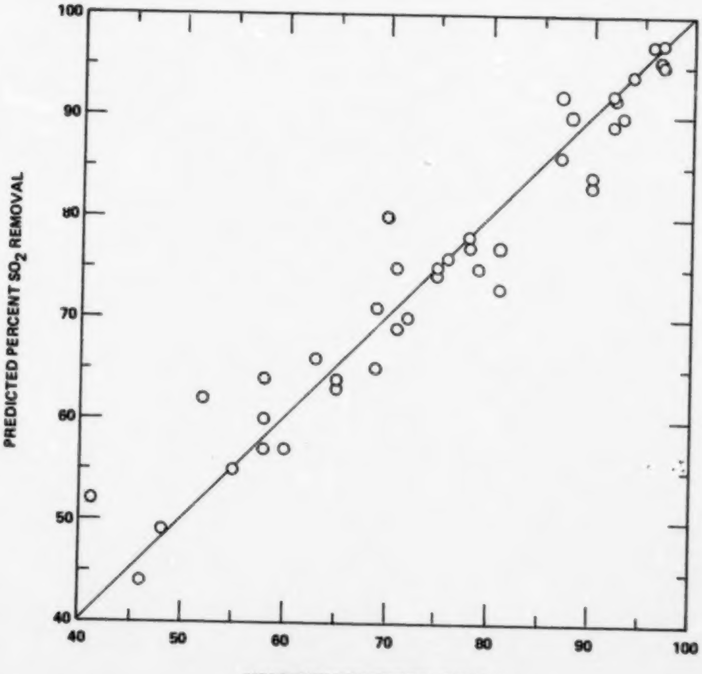


Figure 3-3. Measured vs. Predicted (Eq. 3-2) SO_2 Removal by the Spray Tower

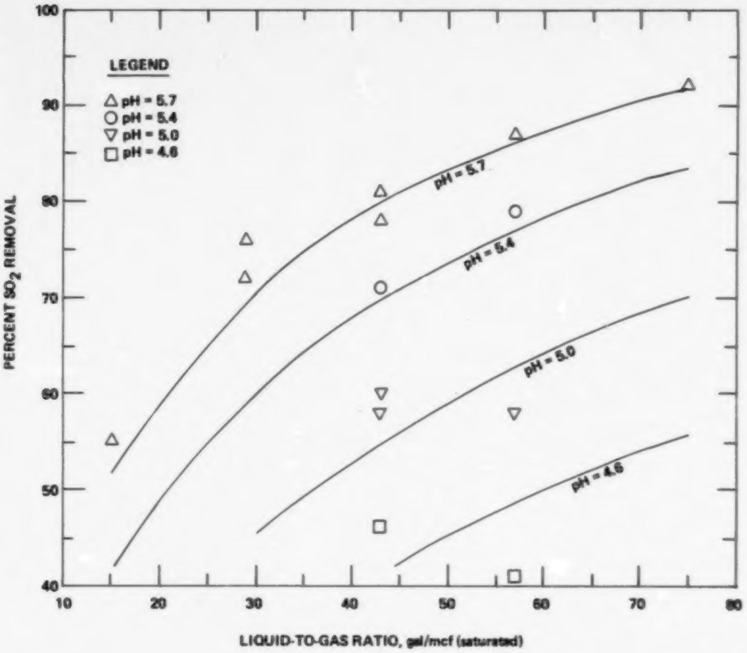


Figure 3-4. Effect of Spray Tower Liquid-to-Gas Ratio and Inlet pH on Spray Tower SO_2 Removal at 600 ppm Adipic Acid

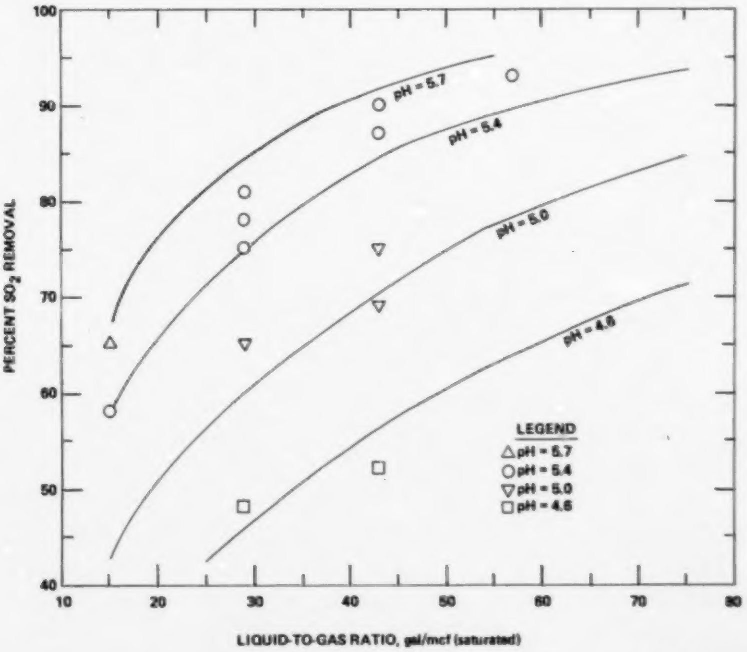


Figure 3-5. Effect of Spray Tower Liquid-to-Gas Ratio and Inlet pH on Spray Tower SO_2 Removal at 1300 ppm Adipic Acid

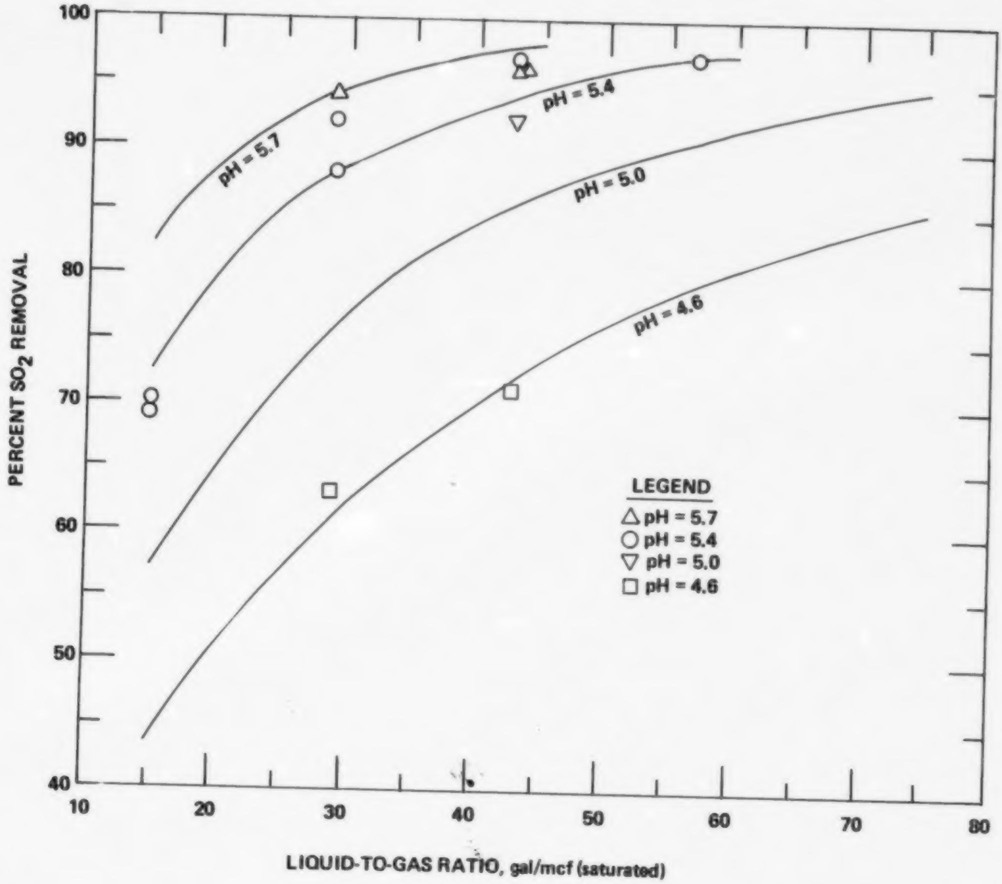


Figure 3-6. Effect of Spray Tower Liquid-to-Gas Ratio and Inlet pH on Spray Tower SO_2 Removal at 2000 ppm Adipic Acid

Thus, within the ranges tested, each 0.4 unit drop in scrubber inlet pH requires a 700 ppm increase in adipic acid concentration to achieve similar percent SO_2 removal.

SO_2 REMOVAL BY VENTURI ALONE AT MINIMUM SLURRY FLOW RATE AND PRESSURE DROP

For a 2 to 4 hour period at the end of each of seven factorial tests with the spray tower alone (Runs VAA210, VAA213, VAA222, VAA231, VAA238, VAA239, and VAA240), the spray tower slurry flow was shut off in order to determine the SO_2 removal achieved by the venturi alone at a minimum slurry flow rate of 125 gpm, minimum pressure drop of 2 to 4 inches H_2O (wide open plug), and 35,000 acfm gas flow rate (venturi L/G = 4.5 gal/mcf). These tests indicated that, at these conditions, the venturi scrubber obtains about 20 percent SO_2 removal at 600 ppm adipic acid concentration and an inlet pH of 5.7, 22 percent SO_2 removal at 1300 ppm adipic acid and a pH of 5.3, and 42 percent SO_2 removal at 2000 ppm adipic acid and a pH of 5.4. Equation 3-2 does not include any corrections for SO_2 removal in the venturi. This should be taken into consideration when using Equation 3-2 with Figures 3-3 through 3-6.

Section 4

EFFECT OF pH ON ADIPIC ACID CONSUMPTION

During both the earlier factorial tests with adipic acid addition (Reference 3) and the latest factorial tests (Section 3), it was noticed that the rate of adipic acid addition required to maintain a desired concentration in the scrubber liquor was substantially reduced when the scrubber inlet pH was controlled at 5.0 or lower. At higher pH operation, it is necessary to add adipic acid at up to about 5 times the theoretical addition rate (as defined below), either because of degradation or decomposition of the adipic acid. Apparently, the degradation or decomposition process is inhibited under low pH conditions.

Although the exact mechanism of adipic acid degradation is still not understood, it was decided to investigate the effect of pH on the adipic acid consumption rate in more detail.

Early in the adipic acid-enhanced lime/limestone testing, it was noted that the SO_2 removal enhancement by the adipic acid is maximized when the scrubber inlet pH is maintained at about 5.2 or higher under the prevailing scrubber conditions (chloride concentrations). This is because most of the adipic acid is ionized and its buffering capacity more fully utilized at these higher inlet pH levels (Reference 7).

Operations at lower pH therefore require higher adipic acid concentrations to maintain the same degree of SO_2 removal efficiency (Section 3), because the ionization and buffer capacity of adipic acid are reduced. However, experience at Shawnee shows that the total adipic acid consumption at a scrubber inlet pH below 5.0 and concentration as high as 4000 ppm is actually lower than at a pH of about 5.4 and 1500 ppm when significant degradation was noted. Potential advantages of low pH operations are obvious:

- Lower operating cost due to lower adipic acid consumption.
- Easier forced oxidation, in-loop or bleed stream, and less air (and compressor energy).
- Essentially complete limestone utilization and improved scrubber operating reliability.
- Reduced sensitivity to limestone type and grind; fine grinding of limestone is probably not required.
- Lower sulfite scaling potential.
- Better prospects (sensitivity) for automatic pH control.
- Greater flexibility for SO_2 emission control; high sensitivity of SO_2 removal to pH allows raising pH to increase the buffer capacity and SO_2 removal when needed.

- Improved acceptance of the concept by plant owners because of the reduced quantity of adipic acid degradation products.
- Applicability to low-sulfur subbituminous and lignite coals containing alkaline ashes which are extractable only at low pH.
- Probable lower cost due to all of the above factors.

Seven runs were conducted on the venturi/spray tower system to investigate the effect of pH on the adipic acid consumption rate. These tests were made with a single effluent hold tank and without forced oxidation. The flow configuration for these tests is the same as that shown in Figure 3-1.

DISCUSSION OF TEST RESULTS

Table 4-1 summarizes the major test conditions and the run-average results for the seven tests made in this series. The scrubber inlet pH and the adipic acid concentration were varied in the tests. All other conditions were held constant.

Theoretical Adipic Acid Consumption Rate. The theoretical adipic acid consumption rate is defined as the rate of adipic acid leaving the scrubber system in the liquor which is entrained in the discharged sludge (filter cake, centrifuge cake, or clarifier underflow) in a closed-liquor-loop operation. The theoretical consumption rate is calculated from the material balances for solids discharged from the scrubber system, solids (or liquor) concentration in the discharged sludge, and adipic acid concentration in the liquor.

Since some adipic acid decomposes to lower-carbon carboxylic acids and the analytical method employed at the Shawnee laboratory determines the total carboxyl group, "adipic acid concentration" as reported throughout this report means "total carboxylic acid expressed as adipic acid." Note that most of the degradation products are also effective as enhancing agents for SO_2 removal.

Effect of pH on Adipic Acid Consumption Rate. As can be seen in Table 4-1, the ratios of actual-to-theoretical adipic acid consumption were all 1.0 at a scrubber inlet pH of 4.60 and 4.85 for Runs 926-1A, 926-1G, and 926-1B, when the adipic acid concentrations were controlled at 4090, 2270, and 1435 ppm, respectively. This indicates that there was essentially no degradation of adipic acid, within the accuracy of the material balance calculations.

Further increase in the scrubber inlet pH to 5.05, 5.25, and 5.50 during Runs 926-1C, 926-1H, and 926-1D resulted in actual-to-theoretical adipic acid consumption ratios of 1.17, 1.24, and 1.59, respectively.

Despite the higher adipic acid concentration required at the lower pH operation, the total adipic acid consumption can be actually less, as can be seen in Table 4-1, in terms of actual adipic acid consumption per ton of SO_2 absorbed.

Table 4-1

RESULTS OF VENTURI/SPRAY TOWER LIMESTONE/ADIPIC ACID TESTS USING A SINGLE TANK WITHOUT FORCED OXIDATION

Major Test Conditions	926-1A	926-1G	926-1B	926-1C	926-1H	926-1D	926-1E
Fly ash loading	High	High	High	High	High	High	High
Gas rate, acfm @ 300°F	35,000	35,000	35,000	35,000	35,000	35,000	35,000
Venturi liquor rate, gpm	600	600	600	600	600	600	600
Spray tower liquor rate, gpm	1600	1600	1600	1600	1600	1600	1600
Percent solids recirculated (controlled)	15	15	15	15	15	15	15
EHT residence time, min	9.1	9.1	9.1	9.1	9.1	9.1	9.1
EHT tank level, ft	8.5	8.5	8.5	8.5	8.5	8.5	8.5
Scrubber inlet pH (controlled)	4.6	4.8	4.8	5.0	5.25	5.5	5.0
Adipic acid concentration, ppm	(1)	(1)	1300	1300	1300	1300	700
Venturi pressure drop, inches H_2O	9	9	9	9	9	9	9
Run-Average Results							
Start-of-run date	12/18/79	1/3/80	1/16/80	1/24/80	1/29/80	2/8/80	2/15/80
Onstream hours	297	244	184	116	169	116	119
Percent SO_2 removal	90	91	84	91	93	96.5	77
Inlet SO_2 concentration, ppm	2650	2250	2115	2150	2150	2410	2450
Adipic acid concentration, ppm	4090	2270	1435	1290	1285	1330	735
Actual adipic acid consumption, lbs/tons SO_2 abs.	--	4.3	3.0	5.7	8.0	9.6	6.0
Adipic acid consumption ratio (actual/theor)	1.0	1.0	1.0	1.17	1.24	1.59	1.0
Percent solids recirculated	15.5	15.1	14.9	15.1	14.9	15.5	15.2
Scrubber inlet pH	4.60	4.85	4.85	5.05	5.25	5.50	5.05
Scrubber outlet pH	4.30	4.50	4.55	4.65	4.85	5.00	4.60
SO_2 make-per-pass, mmole/l	8.05	6.90	6.00	6.60	6.75	7.85	6.35
Limestone utilization, %	97	96	95	95	92	80	95
Scrubber inlet sulfite concentration, ppm	1540	875	965	325	180	135	325
Scrubber outlet sulfite concentration, ppm	1550	1440	1545	695	305	185	710
Sulfite oxidation, %	49	51	49	47	30	17	32
Scrubber inlet gypsum saturation, %	130	116	127	129	118	112	113
Centrifuge cake solids, wt%	69	70	69	66	61	60	60
Mist eliminator restriction, %	--	--	0	--	--	--	0

Notes: (1) Adipic acid concentration controlled at a level to provide 92% SO_2 removal.

Effect of pH and Adipic Acid Concentration on SO_2 Removal. As mentioned in Section 3, the results of factorial tests show that higher adipic acid concentration is required at low pH than at high pH to achieve similar SO_2 removal. This trend is also evident from the results of Runs 926-1A, 926-1G, 926-1C and 926-1H:

	926-1A	926-1G	926-1C	926-1H
Scrubber inlet pH	4.60	4.85	5.05	5.25
Adipic acid conc., ppm	4090	2270	1290	1285
Percent SO_2 removal	90	91	91	93
Inlet SO_2 conc., ppm	2650	2250	2150	2150
Percent limestone utilization	97	96	95	92

Thus, the optimum scrubber inlet pH appears to be 5.0 to 5.1 (Run 926-1C) where adipic acid concentration required is only about 1300 ppm to achieve 91 percent SO_2 removal. More importantly, the adipic acid degradation is insignificant at this pH level (1.17 actual-to-theoretical consumption ratio for Run 926-1C).

Note that SO_2 removal is more sensitive to pH and inlet SO_2 concentrations at the scrubber inlet pH levels of 4.6 to 4.85 tested because the buffer capacity of adipic acid is reduced at the lower pH levels.

Limestone Utilization. One of the benefits of the low pH operation is that very high limestone utilization can be realized. Limestone utilizations were 95 percent or higher at the scrubber inlet pH of 5.05 or lower and 9.1 minutes residence time in the effluent hold tank.

Sulfite Oxidation and Centrifuge Cake Solids. Another important benefit of low pH operation is the ease of forced oxidation of sulfite. A natural oxidation level of about 50 percent was achieved at the scrubber inlet pH of 5.05 or lower, as compared to 15 to 20 percent oxidation at a normal inlet pH of about 5.5. The resulting centrifuge cake solids concentrations were almost 10 percentage points higher for the lower pH operation.

SUMMARY

The following is a summary of the test results:

- Apparent degradation of adipic acid is inhibited at low pH, with or without forced oxidation (see Sections 6 and 7). Without forced oxidition, the critical pH appears to be about 5.0 at the scrubber inlet, below which degradation is minimized (actual-to-theoretical consumption ratio equals 1.0).

- Because of reduced ionization and buffer capacity of acipic acid at low pH, the required adipic acid concentration is 2 to 3 times higher at a scrubber inlet pH of 4.6 to 4.85 than at 5.05 to 5.25 to achieve a similar degree of SO_2 removal (about 91 percent).

- Operation at low pH and high adipic acid concentration results in lower total adipic acid consumption than at high pH and low concentration.

- The optimum scrubber inlet pH for the venturi/spray tower with a single-tank configuration appears to be 5.0 to 5.1 with respect to total adipic acid consumption, limestone utilization, and the sensitivity of SO_2 removal to pH and inlet SO_2 concentration.

- Other benefits obtained when the scrubber inlet pH was held at 5.05 or lower include: high limestone utilization (95 percent or higher), high natural sulfite oxidation (about 50 percent), and the resultant high centrifuge cake solids (near 70 percent).

Section 5

VENTURI SCRUBBER SO_2 REMOVAL WITH LIMESTONE/ADIPIC ACID SLURRY

A series of 12 runs (Runs 927-1A through 927-1L) were made using only the venturi scrubber to determine its maximum SO_2 removal capability with adipic acid-enhanced limestone scrubbing.

While it is recognized that SO_2 removal with the venturi alone would not meet the SO_2 emission standard for high-sulfur coal, even with very high concentrations of adipic acid, scrubbing with the venturi alone could be attractive economically for low-sulfur coal applications where only 70 percent SO_2 removal is required.

A single tank was used without forced oxidation for all tests. The flow configuration for these tests is the same as that shown in Figure 3-1, except the slurry flow to the spray tower (Pumps G-101 and G-204) was turned off.

The slurry flow to the venturi was held constant at 600 gpm for all runs. Variables investigated were adipic acid concentration, gas rate (or venturi liquid-to-gas ratio at a constant slurry flow rate), venturi pressure drop, and inlet pH. Operating conditions common for all runs were:

Fly ash loading: High (3-6 grains/dry scf)
Effluent hold tank level: 8.5 ft
Effluent hold tank residence time: 33.4 minutes
Slurry solids concentration: 15 percent
Solids dewatering equipment: Clarifier and centrifuge

DISCUSSION OF TEST RESULTS

Table 5-1 summarizes the major test conditions and the run-average test results.

Effect of Adipic Acid Concentration. Runs 927-1A, 927-1D, and 927-1E were all operated at a gas rate of 35,000 acfm (@ 300°F), a liquid-to-gas ratio of 21 gal/mcf, a venturi inlet pH of 5.1, and at a pressure drop of about 8.3 inches H_2O. Average SO_2 removal increased from 34 to 41 and 65 percent when the adipic acid concentration was raised from 815 to 1335 and 3985 ppm, respectively. Hourly SO_2 removal data for these three runs are plotted in Figure 5-1. It appears that the SO_2 removal levels off at about 65 percent, suggesting that the overall rate of SO_2 absorption may have been limited by the gas-phase mass transfer above 3500 ppm adipic acid.

Effect of Liquid-to-Gas Ratio. During Runs 927-1B, 927-1C, and 927-1G, the liquid-to-gas ratio was increased to 37 gal/mcf. Average SO_2 removal increased only marginally to 39, 47, and 68 percent, respectively. For these runs, venturi pressure drop was 6 inches H_2O.

Table 5-1

RESULTS OF VENTURI LIMESTONE/ADIPIC ACID TESTS USING A SINGLE TANK WITHOUT FORCED OXIDATION

Major Test Conditions	927-1A	927-1B	927-1J	927-1C	927-1D	927-1E	927-1F	927-1G	927-1H	927-1I	927-1K	927-1L
Fly ash loading	High	High	High	High	High	High	High	High	High	High	High	High
Gas rate, acfm @ 300°F	35,000	20,000	27,500	20,000	35,000	35,000	27,500	20,000	27,500	27,500	27,500	20,000
Venturi liquor rate, gpm	600	600	600	600	600	600	600	600	600	600	600	600
Percent solids recirculated (controlled)	15	15	15	15	15	15	15	15	15	15	15	15
EHT residence time, min	33.4	33.4	33.4	33.4	33.4	33.4	33.4	33.4	33.4	33.4	33.4	33.4
EHT tank level, ft	8.5	8.5	8.5	8.5	8.5	8.5	8.5	8.5	8.5	8.5	8.5	8.5
Venturi inlet liquor pH (controlled)	5.1	5.1	5.1	5.1	5.1	5.1	5.1	5.1	5.1	5.1	4.8	4.8
Adipic acid concentration, ppm	700	700	700	1300	1300	4000	4000	4000	4000	4000	4000	4000
Venturi pressure drop, inches H_2O	6	6	6	6	6	6	6	6	9	12	9	6
Run-Average Results												
Start-of-run date	2/20/80	2/22/80	2/22/80	2/24/80	2/26/80	2/27/80	3/1/80	3/1/80	3/2/80	3/2/80	3/4/80	3/5/80
Onstream hours	43	24	24	48	27	69	26	13	13	21	32	16
Percent SO_2 removal	34	39	33	47	41	65	67(1)	68	69	65	59	62
Inlet SO_2 concentration, ppm	1990	2470	2390	2595	2445	2360	2255	2790	3030	2945	2245	2100
Adipic acid concentration, ppm	815	705	795	1360	1335	3985	3990	4030	4005	4015	4050	4340
Scrubber percent solids recirculated	14.6	14.5	14.6	14.0	16.3	15.4	15.3	14.3	15.0	15.3	14.8	13.6
Scrubber inlet pH	5.05	5.15	5.10	5.15	5.15	5.10	5.10	5.05	5.10	5.10	4.85	4.80
Sulfite concentration in inlet liquor, ppm	365	365	110	330	255	285	460	235	485	555	910	1010
SO_2 Make-per-pass, mmole/l	8.35	6.80	7.65	8.65	12.4	19.0	14.7	13.4	20.3	18.6	12.9	9.20
Limestone utilization, %	92	92	91	91	83	85	85	85	88	91	93	90
Sulfite oxidation, %	36	34	29	32	32	28	23	22	20	32	26	30
Inlet liquor gypsum saturation, %	115	115	135	120	120	125	130	145	125	120	140	125
Centrifuge cake solids, wt%	66	67	68	66	68	61	63	63	63	63	65	68
Mist eliminator restriction, %	--	--	--	--	--	--	--	--	--	--	--	--
Venturi pressure drop, inches H_2O	8.2	5.9	5.9	6.0	8.4	8.3	5.9	6.0	8.7	11.1	8.8	6.2

Note: (1) SO_2 removal dropped to 60% when inlet SO_2 concentration increased to 2870 ppm under replicate conditions.

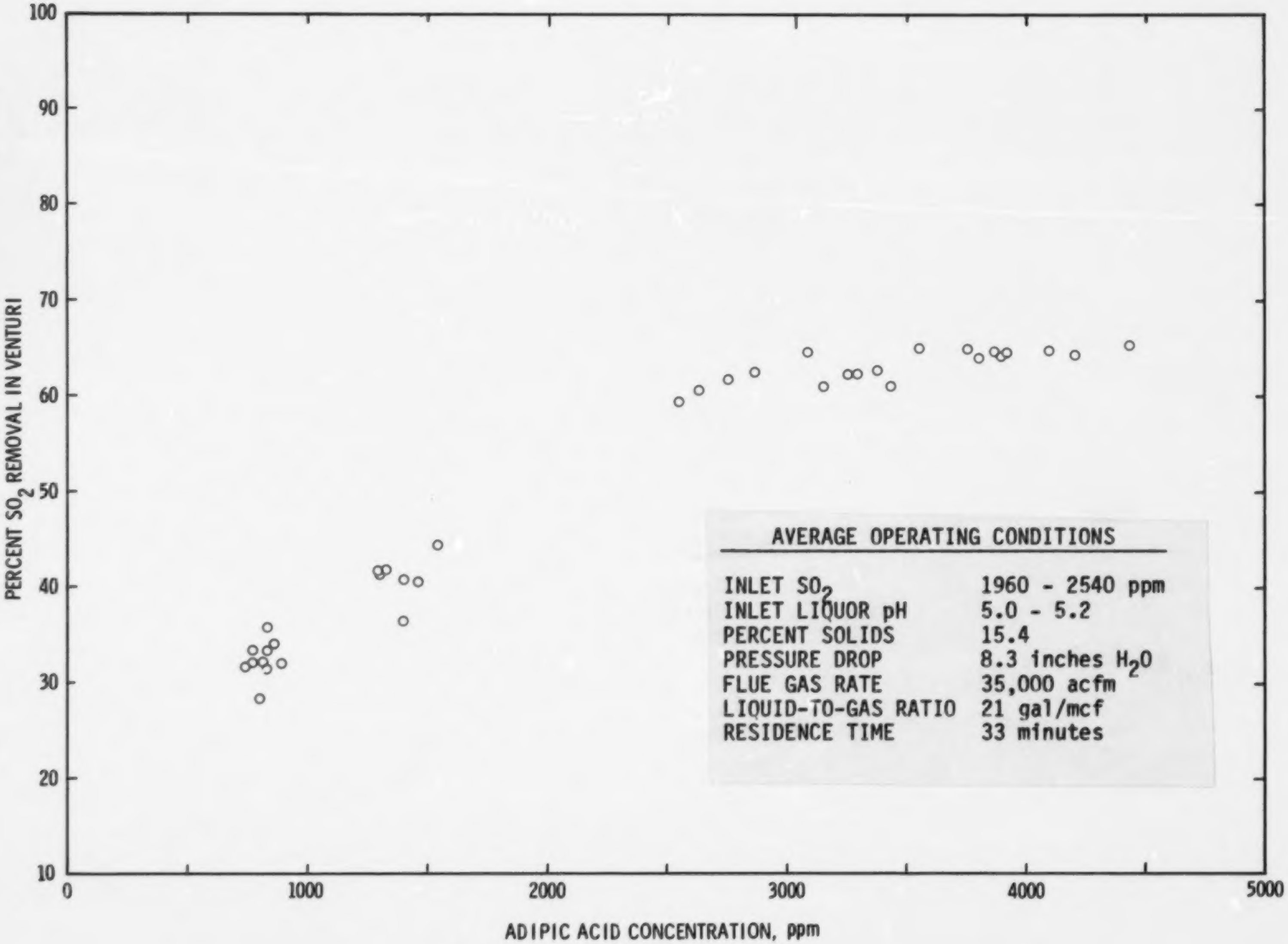


Figure 5-1. Percent SO_2 Removal vs. Adipic Acid Concentration During Limestone Runs 927-1A, 1D and 1E

Although SO_2 removal was below 70 percent with high inlet SO_2 concentration, the venturi-only mode of operation with limestone/adipic acid slurry may be viable for low-sulfur coal applications where inlet SO_2 concentrations are less than 1000 ppm.

Effect of Venturi Pressure Drop. During Runs 927-1F, 927-1H, and 927-1I, the venturi pressure drop was varied at 5.9, 8.7, and 11.1 inches H_2O, respectively. For these runs, adipic acid concentration was maintained at 4000 ppm, liquid-to-gas ratio was controlled at 27 gal/mcf, and the inlet pH was controlled at 5.1. SO_2 removal was 60 percent at 5.9 inches H_2O pressure drop, and appeared to level off at 65 to 69 percent at 8.7 and 11.1 inches H_2O.

Effect of Venturi Inlet pH. Run 927-1K was made under the same conditions as Run 927-1H, except for the scrubber inlet pH. SO_2 removal increased significantly from 59 percent at 2245 ppm inlet SO_2 concentration and at 4.85 inlet pH to 69 percent at 3030 ppm inlet SO_2 concentration and at 5.10 inlet pH. Similar sensitivity of SO_2 removal to pH can be observed by comparing Runs 927-1G and 927-1L.

SUMMARY

Based on the test results, the following conclusions can be made:

- At a liquid-to-gas ratio of 21 gal/mcf, a venturi inlet pH of 5.1, and a venturi pressure drop of 8.3 inches H_2O, SO_2 removal appears to level off at 65 percent above 3500 ppm adipic acid. (SO_2 removals greater than 65 percent may be possible at pH higher than 5.1.)
- Increasing the liquid-to-gas ratio to 37 gal/mcf (with a somewhat reduced pressure drop of 6 inches H_2O) improves SO_2 removal marginally.
- With low-sulfur coals producing less than 1000 ppm inlet SO_2 concentration, 70 percent SO_2 removal should be achievable at 5.1 inlet pH, 4000 ppm adipic acid, 6 to 8 inches H_2O pressure drop, and 21-37 gal/mcf liquid-to-gas ratio.
- SO_2 removal is sensitive to inlet pH (4.8 to 5.1) and adipic acid concentration (700 to 3500 ppm), but is insensitive to liquid-to-gas ratio (21 to 37 gal/mcf) and venturi pressure drop (6 to 11 inches H_2O).

Section 6

LIMESTONE/ADIPIC ACID TESTING ON THE VENTURI/SPRAY TOWER WITH ONE TANK AND FORCED OXIDATION

Following the venturi-only testing, the venturi/spray tower system was modified to allow testing in a single-tank forced-oxidation mode. Seven runs were made, including four runs with only the venturi.

Although sulfite oxidation of 99 percent or higher was achieved for the runs with forced oxidation, limestone blinding was encountered as evidenced by poor limestone utilization. The long (50 ft) crossover line which routed the venturi and spray tower effluent slurries to the oxidation tank apparently behaved as an effective plug-flow reactor in which calcium sulfite precipitated preferentially on the alkaline limestone particles in the effluent slurry deficient in calcium sulfite solid crystal seeds.

SYSTEM DESCRIPTION

Figure 6-1 is a schematic flow diagram of the venturi/spray tower system using a single tank (D-208) in which compressed air is injected through a 3-inch diameter open-ended pipe ell. The venturi and spray tower effluent slurries are routed to the oxidation tank via a 16-inch diameter crossover line about 50 ft long. This crossover line is operated full (490 gallons) of slurry because nearly its entire length is below the oxidation tank liquid level. The line acts as a plug-flow reactor as previously mentioned. It is emphasized that this setup is necessitated by the limited availability of space and is unique to the Shawnee Test Facility.

A severe cavitation problem in the slurry recirculation pumps during initial startup was solved by installing a baffle near the pump suction nozzles and by moving the air injection point higher, to between the two agitator turbines. Both turbines propel the slurry downward. Figure 6-2 shows the arrangement of the modified oxidation tank.

DISCUSSION OF TEST RESULTS

Tables 6-1 and 6-2 summarize the results of the single-tank forced oxidation tests with both the venturi and the spray tower in operation, and with the venturi alone, respectively. The initial test plan called for variations of the scrubber inlet pH and adipic acid concentration, to observe the effects on adipic acid consumption under forced oxidation conditions (to compare with the results presented in Section 4 without forced oxidation). However, the original test objectives were modified in favor of a more thorough study of the limestone blinding phenomenon when it was encountered.

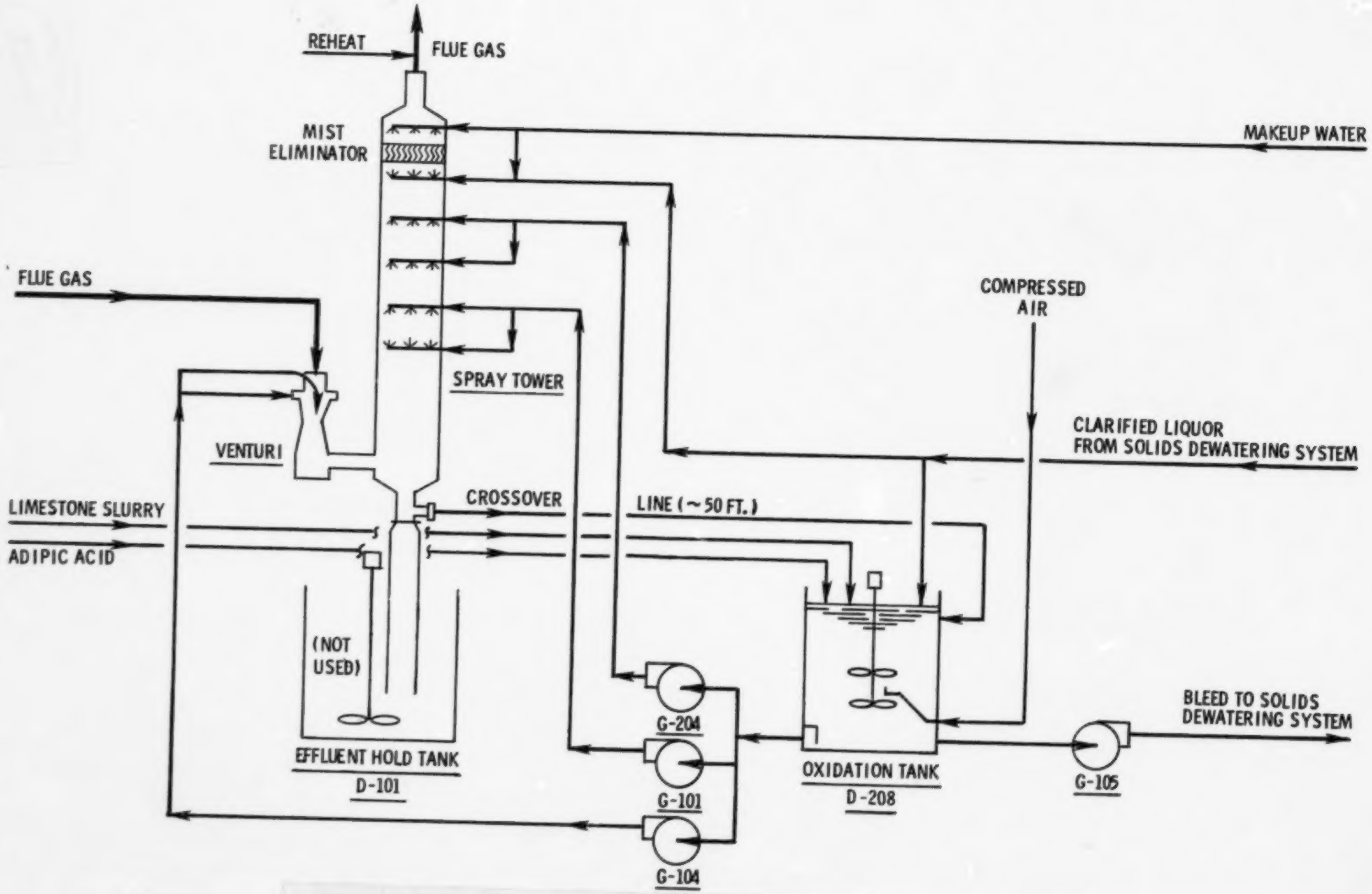
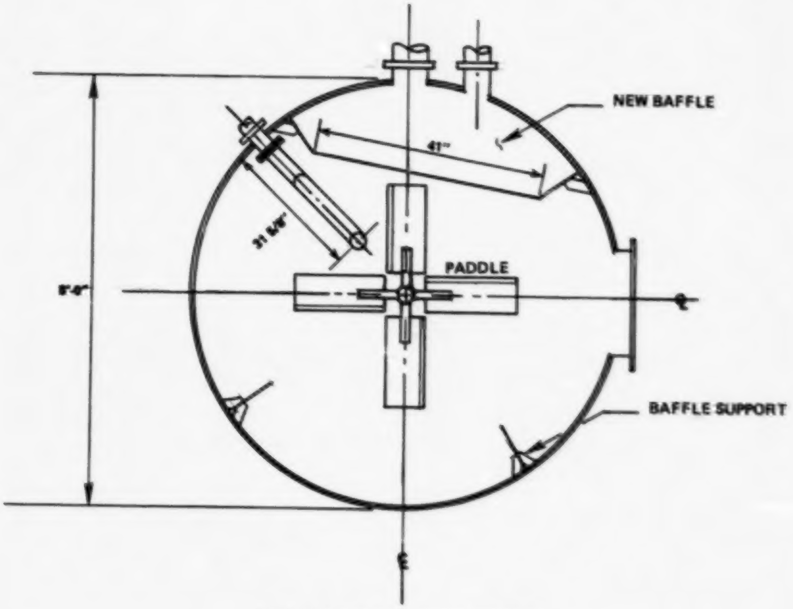
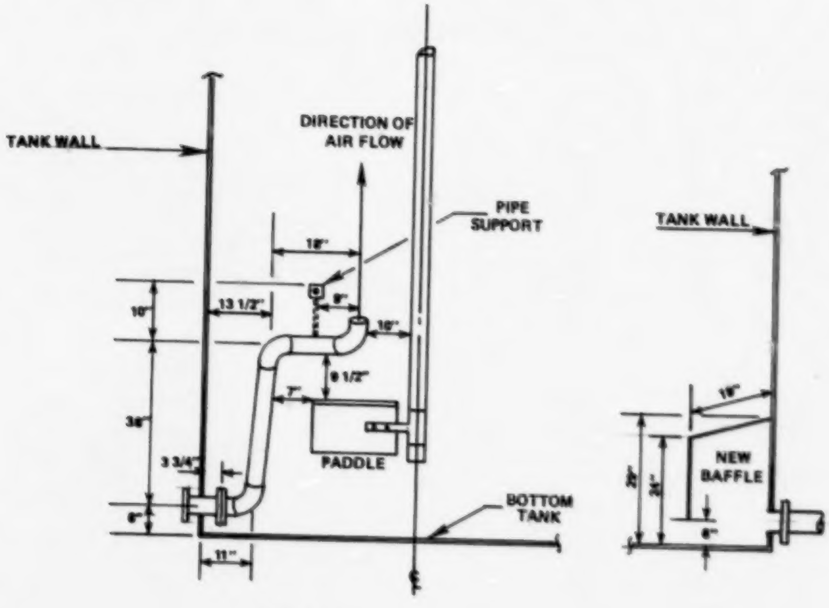

REHEAT
FLUE GAS
MIST ELIMINATOR
MAKEUP WATER
FLUE GAS
COMPRESSED AIR
SPRAY TOWER
CLARIFIED LIQUOR FROM SOLIDS DEWATERING SYSTEM
VENTURI
LIMESTONE SLURRY
ADIPIC ACID
CROSSOVER
LINE (~ 50 FT.)
(NOT USED)
G-204
G-101
G-104
EFFLUENT HOLD TANK
D-101
OXIDATION TANK
D-208
BLEED TO SOLIDS DEWATERING SYSTEM
G-105

Figure 6-1. Flow Diagram for the Venturi/Spray Tower System With One Tank and Forced Oxidation


NEW BAFFLE
41"
31 5/8"
PADDLE
8'-0"
℄
BAFFLE SUPPORT
℄

OVERHEAD VIEW OF TANK BOTTOM


TANK WALL
DIRECTION OF AIR FLOW
PIPE SUPPORT
18"
10"
13 1/2"
8"
10"
9 1/2"
7"
36"
3 3/4"
PADDLE
6"
BOTTOM TANK
11"
℄
TANK WALL
18"
29"
24"
NEW BAFFLE
6"

SIDE VIEW

SIDE VIEW

Figure 6-2. Arrangement of Modified Venturi/Spray Tower Oxidation Tank (D-208)

Table 6-1

RESULTS OF VENTURI/SPRAY TOWER LIMESTONE/ADIPIC ACID TESTS WITH ONE TANK AND FORCED OXIDATION

Major Test Conditions	914-1A	914-1B	914-1C
Fly ash loading	High	High	High
Gas rate, acfm @ 300°F	35,000	35,000	35,000
Venturi liquor rate, gpm	600	600	600
Spray tower liquor rate, gpm	1600	1600	1600
Percent solids recirculated (controlled)	15	15	15
Oxidation tank residence time, min	2.9	2.9	2.9
Oxidation tank level, ft	17	17	17
Scrubber inlet pH (controlled)	4.6	5.1	4.6
Adipic acid concentration, ppm	4000	4000	4000
Air rate to oxidizer, scfm	200	200/300	0
Venturi pressure drop, inches H_2O	9	9	9
Run-Average Results			
Start-of-run date	3/13/80	3/19/80	4/7/80
Onstream hours	105	11	47
Percent SO_2 removal	91.6	(1)	92.6
Inlet SO_2 concentration, ppm	1980		1955
Adipic acid concentration, ppm	4040		4225
Adipic acid consumption ratio (actual/theor.)	3.41		1.59
Actual adipic acid consumption, lbs/ton SO_2 absorbed	64.1		43.3
Percent solids recirculated	15.9		14.9
Scrubber inlet pH	4.60		4.66
Sulfite concentration in inlet liquor, ppm	1250		862
SO_2 make-per-pass, mmole/l	6.1		6.1
Limestone utilization, %	46		93
Sulfite oxidation, %	98.7		32
Gypsum saturation in inlet liquor, %	145		120
Centrifuge cake solids, wt%	79		65
Air stoichiometry, atom O/mole SO_2 abs.	1.9		0
Mist eliminator restriction, %	--		3

(1) No steady state was established due to severe limestone blinding.

Table 6-2

RESULTS OF VENTURI LIMESTONE/ADIPIC ACID TESTS WITH ONE TANK AND FORCED OXIDATION

Major Test Conditions	927-1M	927-1N	927-1O	927-1P
Fly ash loading	High	High	High	High
Gas rate, acfm @ 300°F	30,000	20,000	20,000	30,000
Venturi liquor rate, gpm	600	600	600	600
Percent solids recirculated (controlled)	15	15	15	15
Oxidation tank residence time, min	10.6	10.6	10.6	10.6
Oxidation tank level, ft	17	17	17	17
Venturi inlet liquor pH (controlled)	5.0	5.0	--	5.0
Venturi inlet liquor limestone stoichiometry (controlled)	--	--	1.2	--
Adipic acid concentration, ppm	4000	4000	4000	4000
Air rate to oxidizer, scfm	300	300	300	0
Venturi pressure drop, inches H_2O	9	9(1)	9(1)	9
Run-Average Results				
Start-of-run date	3/21/80	3/26/80	3/30/80	4/2/80
Onstream hours	115	13	75	108
Percent SO_2 removal	71.5	77.4	67.4	69.6
Inlet SO_2 concentration, ppm	2260	2030	2070	2225
Adipic acid concentration, ppm	4170	3960	4130	3960
Adipic acid consumption ratio (actual/theor.)	2.19	3.0	1.93	2.26
Actual adipic acid consumption, lbs/ton SO_2 absorbed	32.0	50.8	28.6	62.5
Percent solids recirculated	15.0	16.1	15.0	14.9
Scrubber inlet pH	5.05	5.15	4.55	5.07
Sulfite concentration in inlet liquor, ppm	75	44	28	349
SO_2 make-per-pass, mmole/l	17.1	11.1	9.9	16.4
Limestone utilization, %	50	35	85	54
Sulfite oxidation, %	99.4	99.2	99.2	23
Gypsum saturation in inlet liquor, %	105	105	110	125
Centrifuge cake solids, wt%	78	79	78	63
Air stoichiometry, atom O/mole SO_2 abs.	3.8	5.9	6.6	0
Mist eliminator restriction, %	--	--	--	--

(1) Actual pressure drop was about 7 inches H_2O because of a problem with the adjustable plug mechanism and low gas flow rate.

Initial Tests. In Run 914-1A, a total slurry flow rate of 2200 gpm resulted in 2.9 minutes residence time in the oxidation tank (Table 6-1), 98.7 percent sulfite oxidation in the solids, high inlet liquor sulfite concentration (1250 ppm), and poor limestone utilization of 46 percent despite a low scrubber inlet pH of 4.6.

To reduce the high inlet liquor sulfite concentration, Run 914-1B was first run at higher pH (5.1 vs. 4.6) and then at increased oxidation intensity (air rate 300 scfm vs. 200 scfm). However, no indication of increased limestone utilization was noted and the run was terminated.

Venturi-Only Test. The low oxidation tank residence time of 2.9 minutes was increased to 10.6 minutes during Runs 927-1M, 927-1N, and 927-10 (Table 6-2) by operation of the venturi only (600 gpm). This necessarily raised the SO_2 make-per-pass to 17.1 m-moles/liter (Run 927-1M) which was reduced to 11.1 m-moles/liter in Run 927-1N. The limestone utilization was still low and a run at a controlled limestone stoichiometry of 1.2 (Run 927-10) confirmed that limestone blinding was occuring in the crossover line because the scrubber inlet pH of 4.55 was lower than expected. This line is in effect a 50 second residence time plug-flow reactor to which is fed slurry depleted in calcium sulfite seed crystals and in which a favorable environment is provided for the liquor sulfite to precipitate on limestone particles before reaching the oxidation tank.

Base Case Tests Without Forced Oxidation. Run 927-1P was made under the same conditions as Run 927-1M except that the air to the oxidizer was shut off to provide a base case run without forced oxidation. Limestone utilization remained poor (54 percent) due to the combined effect of continued high SO_2 make-per-pass (16.4 m-moles/liter) and long residence time (near 50 seconds) in the crossover line.

Run 914-1C was made under the same conditions as Run 914-1A except without forced oxidation. With an SO_2 make-per-pass of only 6.1 m-moles/liter and 13 seconds residence time in the crossover line, combined with sufficient calcium sulfite solid crystal seeds (32 percent oxidation), limestone utilization improved to 93 percent.

Effect of pH and Limestone Utilization on Adipic Acid Consumption. Section 4 mentioned that essentially no degradation of acipic acid occurs at a scrubber inlet pH below 5.0 when oxidation is not forced. In these tests adipic acid degradation appeared to increase with forced oxidation. In addition, it was observed that poor limestone utilization increases the degradation. These observations are more clearly seen in the following table:

	927-1M	927-1N	927-10
Venturi inlet pH	5.05	5.15	4.55
Percent limestone utilization	50	35	85
Adipic acid consumption ratio (Actual/Theoretical)	2.19	3.0	1.93
Percent unaccounted loss of adipic acid	54.3	66.7	48.2
Actual adipic acid consumption, lbs/ton SO_2 absorbed	32.0	50.8	28.6
Unaccounted adipic acid loss, lbs/ton SO_2 absorbed	17.4	33.9	13.8

SUMMARY

The following is a summary of the test results and findings:

- Good sulfite oxidation of 99 percent or higher was achieved in the solids despite poor limestone utilization.
- Limestone blinding occurred in the 50 ft long crossover line which transfers the venturi and spray tower effluent slurries to the oxidation tank and which behaved as an effective plug-flow reactor for calcium sulfite precipitation. This peculiarity in flow configuration is unique to the Shawnee Test Facility.
- Limestone blinding caused by the long crossover line and high SO_2 make-per-pass could not be prevented by increasing the oxidation intensity in the oxidation tank to reduce the sulfite concentration in the scrubber inlet liquor, even at SO_2 make-per-pass values as low as 6.1 m-moles/liter, and was compounded by depletion of calcium sulfite seed crystals in the scrubber effluent.
- Actual-to-theoretical adipic acid consumption ratio and total actual adipic acid requirement (lbs per ton SO_2 absorbed) increase with forced oxidation, increasing pH, and decreasing limestone utilization.

Section 7

LIMESTONE/ADIPIC ACID TESTING ON THE VENTURI/SPRAY TOWER WITH TWO TANKS AND FORCED OXIDATION

Operation with two tanks in series, with forced oxidation in the first tank and limestone added to the second tank, has several advantages over the single-tank operation with forced oxidation:

- Low pH (scrubber-effluent pH) in the first tank (oxidation tank) promotes sulfite oxidation.
- The possibility of limestone blinding by calcium sulfite is decreased because fresh limestone is added after the oxidation tank.
- Limestone utilization is increased with two tanks in series which approximate a plug-flow reactor for limestone dissolution.
- Extra residence time for calcium sulfate crystallization is provided by the second tank.
- The second tank provides air-free suction for the slurry recirculation and bleed pumps, thus avoiding pump cavitation.

Earlier test results from the TCA system with limestone/adipic acid and forced oxidation have shown two-tank operation to be superior to the single-tank mode (Reference 3). Eight runs (Runs 916-1A through 916-1H) were made to confirm this conclusion on the venturi/spray tower system using two tanks in series. A schematic flow diagram is shown in Figure 7-1. Air is injected into the first tank (D-208) while limestone and adipic acid are added to the second tank (D-101). The detailed arrangement of the oxidation tank (8 ft diameter) is shown in Figure 6-2.

DISCUSSION OF TEST RESULTS

Table 7-1 summarizes the results of the eight runs made with two tanks in series, including one run (Run 916-1H) without forced oxidation. In general, good SO_2 removal was achieved with excellent oxidation of the solids for all the forced oxidation tests. However, as in the tests with forced oxidation using a single tank (Section 6), calcium sulfite blinding of limestone in the crossover line continued to reduce the limestone utilization below the level normally expected with two-tank operation. This remained true despite the efforts to increase limestone utilization by either increasing the oxidation intensity or lowering the SO_2 make-per-pass.

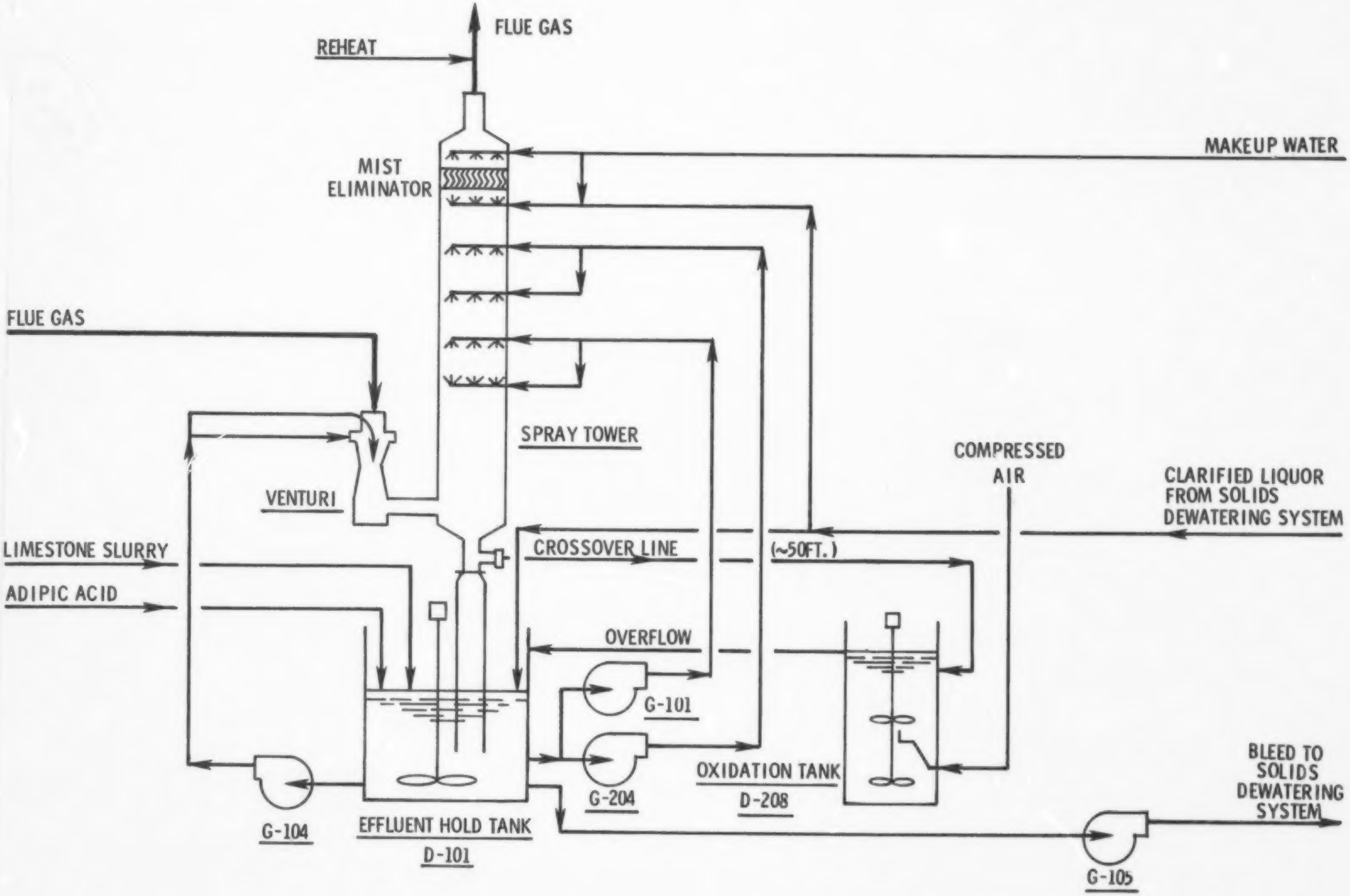

FLUE GAS
REHEAT
MIST ELIMINATOR
MAKEUP WATER
FLUE GAS
SPRAY TOWER
VENTURI
COMPRESSED AIR
CLARIFIED LIQUOR FROM SOLIDS DEWATERING SYSTEM
LIMESTONE SLURRY
CROSSOVER LINE
(~50FT.)
ADIPIC ACID
OVERFLOW
G-101
G-204
OXIDATION TANK
D-208
BLEED TO SOLIDS DEWATERING SYSTEM
G-104
EFFLUENT HOLD TANK
D-101
G-105

Figure 7-1. Flow Diagram for the Venturi/Spray Tower System With Two Tanks and Forced Oxidation

Table 7-1

RESULTS OF VENTURI/SPRAY TOWER LIMESTONE/ADIPIC ACID TESTS WITH TWO TANKS AND FORCED OXIDATION

Major Test Conditions	916-1A	916-1B	916-1C	916-1D	916-1E	916-1F	916-1G	916-1H
Fly ash loading	High	High	High	High	High	High	High	High
Flue gas rate, acfm @ 300°F	35,000	30,000	30,000	30,000	20,000	30,000	30,000	30,000
Venturi liquor rate, gpm	600	600	600	600	600	600	600	600
Spray tower liquor rate, gpm	500	1200	1200	1200	1600	1200	1200	1200
Percent solids recirculated (controlled)	15	15	15	15	15	15	15	15
EHT res. time (min)/tank level (ft)	18.2/8.5	11.1/8.5	11.1/8.5	11.1/8.5	9.1/8.5	11.1/8.5	11.1/8.5	11.1/8.5
Oxid. tk. res. time (min)/tank level (ft)	6.1/18	3.8/18	3.8/18	3.8/18	3.1/18	3.8/18	3.8/18	3.8/18
Inlet liquor pH (controlled)	4.8	4.8	4.8	5.1	5.1	5.4	5.4	5.1
Adipic acid concentration, ppm	4000	4000	4000/1500	1500	1500	1500	1500	1500
Air stoich. to oxidizer, atom O/mole SO_2 absorbed	1.5	1.5	1.5	1.5	1.5	1.5	2.5	0
Venturi pressure drop, inches H_2O (1)	9	6	6	6	7	6	6	6
Run-Average Results								
Start-of-run date	4/24/80	4/28/80	5/2/80	5/6/80	5/13/80	5/19/80	5/22/80	5/24/80
Onstream hours	89	77	92	163	145	71	61	84
Percent SO_2 removal	88.1	96.5	93.7 → 91.2	92.4	98.0	89.2	93.7	85.5
Inlet SO_2 concentration, ppm	2010	2170	2330	2220	1880	2260	2150	2500
Adipic acid concentration, ppm	3990	4015	3450/1600	1490	1540	1510	1550	1440
Adipic acid consumption ratio, (actual/theor)	2.84	1.37	--	--	4.89	2.30	3.32	2.03
Actual adipic acid consumption, lbs/ton SO_2 abs.	32.7	15.8	--	--	8.91	7.75	8.76	11.4
Percent solids recirculated	15.7	15.3	15.1	15.0	14.9	15.9	15.6	15.6
Scrubber inlet pH	4.77	4.83	4.86	5.12	5.13	5.14	5.33	5.12
Oxidation tank pH	4.52	4.70	4.67	5.03	5.09	5.06	5.24	4.77
Limestone utilization, %	70	84	88	72	86	46	61	96
Sulfite oxidation, %	99.4	99.7	99.7	99.8	99.5	99.4	99.6	50.4
Inlet liquor gypsum saturation, %	120	105	110	105	105	145	120	120
Sulfite conc. in inlet liquor, ppm	380	71	137	81	60	583	33	317
Avg. air flow rate, scfm	151	158	158	150	90	148	246	0
Centrifuge cake solids, %	79	76	34(2)	84(3)	88(3)	84(3)	85(3)	59(3)
Mist eliminator restriction, %	--	1	--	--	--	--	--	1
SO_2 make-per-pass, m-mole/liter	12.0	7.4	7.5-7.7	7.3	3.6	7.1	7.1	7.6

(1) Venturi operated with plug wide open for all runs except for Run 916-1A where pressure drop was controlled at 9 inches H_2O.

(2) System operated with clarifier only.

(3) Drum filter used in place of centrifuge.

Forced Oxidation Testing. During the testing covered by Runs 916-1A through 916-1G, several measures were taken to eliminate or minimize the effect of the crossover line. Operating parameters explored included:

- Liquid-to-gas ratios in the spray tower of 17.8 to 100 gal/mcf
- SO_2 make-per-pass of 3.6 to 12.0 m-moles/liter
- Adipic acid concentrations of 1490-4015 ppm
- Scrubber inlet pH of 4.77 to 5.33
- Air stoichiometry to the oxidizer of 1.5 to 2.5 atoms O/mole SO_2 absorbed

However, the overriding tendency of the crossover line to act as a plug-flow reactor, as described in Section 6, could not be eliminated.

Base Case Test Without Forced Oxidation. Run 916-1H was made under the same conditions as Run 916-1D except that the air to the oxidizer was turned off. Significantly, the limestone utilization improved to 96 percent because sufficient calcium sulfite solid seeds were available (50.4 percent oxidation) and blinding of limestone by calcium sulfite in the crossover line was eliminated.

SUMMARY

The following is a summary of the test results:

- Good SO_2 removal and excellent sulfite oxidation (99.4 to 99.8 percent) were achieved with the two-tank forced oxidation system.
- Limestone utilization for the two-tank operation was higher than for single-tank operation (Section 6) but below that expected with two-tank operation without limestone blinding.
- As in the single-tank operation with forced oxidation (Section 6), limestone blinding caused by the crossover line and high SO_2 make-per-pass cannot be eliminated by increasing the oxidation intensity to reduce sulfite concentration in the scrubber inlet liquor.
- Reducing the SO_2 make-per-pass (Run 916-1E), and hence the scrubber effluent sulfite concentration, improved limestone utilization but not to the expected level.

Section 8

BLEED STREAM OXIDATION OF LIMESTONE/ADIPIC ACID SLURRY FROM THE VENTURI/SPRAY TOWER SYSTEM

In April 1979, prior to this reporting period, five bleed stream oxidation tests were made on the venturi/spray tower system using limestone slurry with 1500 ppm of adipic acid (Reference 3). At that time, good sulfite oxidation of 99 percent was achieved when the slurry pH in the oxidation tank was kept below about 6.0 by recycling 60 gpm of oxidation tank slurry back to the effluent hold tank. Satisfactory oxidation (95 percent) was also obtained without the recycle, but at the high oxidation tank residence time of about 7.5 hours for the bleed stream.

Recent tests with adipic acid additive have demonstrated several advantages of operating at low pH and high adipic acid concentration (see Section 4). Therefore, three tests (Runs 915-1A, 915-1B, and 915-1C) were conducted in April 1980 to see if operating at reduced pH was conducive to bleed stream oxidation. The flow diagram of the bleed stream oxidation tests on the venturi/spray tower system is shown in Figure 8-1. The same oxidation tank used in one-tank and two-tank in-loop forced oxidation (Sections 6 and 7) was used in these three tests. The detailed arrangement of the oxidation tank is shown in Figure 6-2.

DISCUSSION OF TEST RESULTS

The results of bleed stream oxidation tests at low pH are given in Table 8-1. All tests achieved better than 95 percent SO_2 removal at 4.8 to 5.1 scrubber inlet pH and about 4000 ppm adipic acid. Average limestone utilizations were 88 to 91 percent.

Good sulfite oxidation of 98 percent was achieved only in Run 915-1C when the scrubber inlet pH was controlled at 4.8 with an air stoichiometry of 1.80 atoms oxygen/mole SO_2 absorbed. Oxidation was only about 70 percent at 5.0 scrubber inlet pH and 1.55 air stoichiometry (Run 915-1A), or at 5.1 scrubber inlet pH and 2.10 air stoichiometry (Run 915-1B).

The oxidation tank pH was 5.4, 5.7, and 4.8 for Runs 915-1A, 915-1B, and 915-1C, respectively, as compared with 5.5 to 5.6 for runs made earlier in April 1979 when good oxidation was achieved at 1.50 to 1.85 air stoichiometry. The lower oxidation efficiency for the recent tests may be attributed to the poor oxidizer arrangement shown in Figure 6-2.

As has been observed previously, adipic acid degradation increased with pH during these runs. For Runs 915-1C, 915-1A, and 915-1B, under similar limestone utilizations, the actual-to-theoretical adipic acid consumption ratios were 1.26, 3.33, and 5.20, respectively, when the scrubber inlet pH increased from 4.8 to 5.0 and to 5.1, and the oxidation tank pH increased concurrently from 4.8 to 5.4 and to 5.7. Actual adipic acid consumption increased from 15.4 to 40.1 and to 44.5 lbs/ton SO_2 absorbed, respectively.

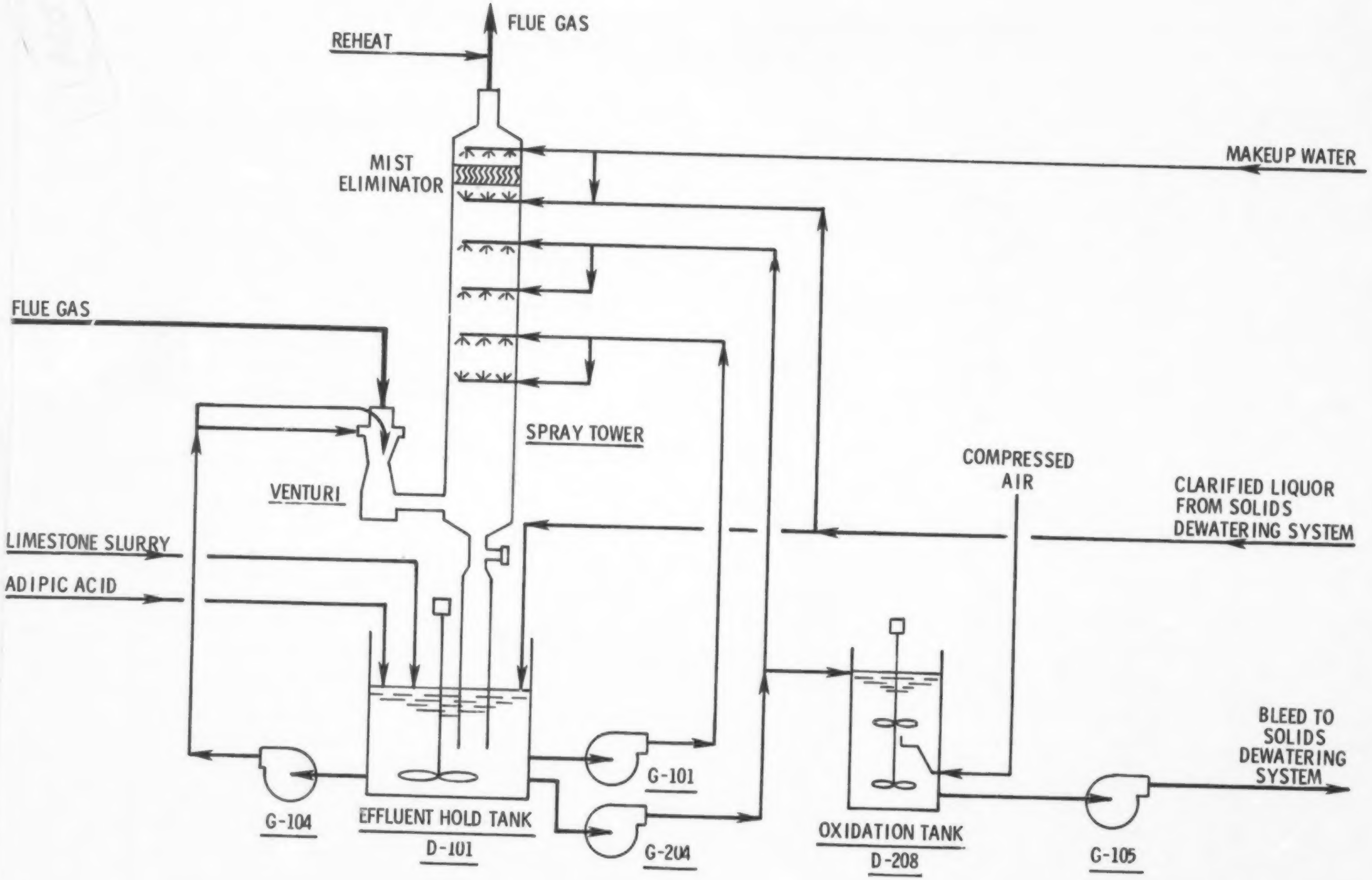

FLUE GAS
REHEAT
MIST ELIMINATOR
MAKEUP WATER
FLUE GAS
SPRAY TOWER
VENTURI
COMPRESSED AIR
CLARIFIED LIQUOR FROM SOLIDS DEWATERING SYSTEM
LIMESTONE SLURRY
ADIPIC ACID
BLEED TO SOLIDS DEWATERING SYSTEM
G-101
G-104
EFFLUENT HOLD TANK
D-101
G-204
OXIDATION TANK
D-208
G-105

Figure 8-1. Flow Diagram for Bleed Stream Oxidation in the Venturi/Spray Tower System

Table 8-1

RESULTS OF VENTURI/SPRAY TOWER LIMESTONE/ADIPIC ACID TESTS WITH BLEED STREAM OXIDATION

Major Test Conditions	915-1A	915-1B	915-1C
Fly ash loading	High	High	High
Flue gas rate, acfm @ 300°F	35,000	35,000	35,000
Venturi liquor rate, gpm	600	600	600
Spray tower liquor rate, gpm	1600	1600	1600
Percent solids recirculated (controlled)	15	15	15
EHT Res. time (min)/tank level (ft)	9.1/8.5	9.1/8.5	9.1/8.5
Oxid. Tk. Res. time (min)/tank level (ft)	-/17	-/17	-/17
Scrubber inlet pH (controlled)	5.1	5.1	4.8
Adipic acid concentration, ppm	4000	4000	4000
Air rate to oxidizer, scfm	200	300	200
Venturi pressure drop, inches H_2O	9	9	9
Run-Average Results			
Start-of-run date	4/10/80	4/14/80	4/15/80
Onstream hours	98	24	127
Percent SO_2 removal	97.6	98.0	96.0
Inlet SO_2 concentration, ppm	2340	2550	2030
Adipic acid concentration, ppm	3840	4045	4140
Adipic acid consumption ratio, (actual/theor.)	3.33	5.20	1.26
Actual adipic acid consumption, lbs/ton SO_2 absorbed	40.1	44.5	15.4
Percent solids recirculated	15.2	15.5	15.6
Scrubber inlet pH	4.99	5.09	4.82
Oxidation tank pH	5.40	5.70	4.80
Limestone utilization, %	91	90	88
Sulfite oxidation in oxidation tank, %	69	73	98
Sulfite oxidation in scrubber inlet, %	26	25	54
Gypsum sat'n. in oxidation tank, %	105	105	100
Gypsum sat'n. in scrubber inlet, %	120	115	105
Oxidation tank liquor $SO_3^=$ concentration, ppm	115	95	140
Air stoich., lb atoms O/lb mole SO_2 absorbed	1.55	2.10	1.80
Centrifuge cake solids, wt%	70	79	79
Mist eliminator restriction, %	--	--	3

Previous Shawnee data indicated that the dewatering properties of slurries from bleed stream oxidation are better than those of unoxidized slurries but inferior to those from in-loop forced oxidation. For Run 915-1C, with 98 percent sulfite oxidation and 4140 ppm adipic acid, the initial settling rate of solids averaged only 0.3 cm/min, somewhat better than the 0.2 cm/min settling rate for unoxidized slurry (see Section 10). For the bleed stream oxidation runs made in April 1979, the average settling rate was much higher at 0.8 cm/min for slurries with good oxidation (95 percent or higher) and with lower 1500 ppm adipic acid concentration. These values for bleed stream oxidation are in the lower range of 0.3 to 1.6 cm/min reported in Table 10-1 for all the oxidized limestone slurry with adipic acid.

SUMMARY

At a scrubber inlet pH of 4.8 and about 4000 ppm adipic acid concentration, 98 percent oxidation of sulfite was achieved in the bleed stream oxidation tank (4.8 pH) with an air stoichiometry of 1.8 atoms oxygen/mole SO_2 absorbed. The SO_2 removal was 96 percent at 2030 ppm inlet SO_2 concentration and the limestone utilization was 88 percent. The actual-to-theoretical adipic acid consumption ratio was 1.26 and the actual adipic acid consumption was 15.4 lbs/ton SO_2 absorbed (8.7 lbs/ton limestone fed).

Section 9

LIMESTONE BLINDING BY CALCIUM SULFITE

Blinding of limestone as evidenced by low limestone utilization has been encountered during limestone tests with and without adipic acid enhancement.

The limestone blinding is most common under in-loop forced oxidation conditions, where the recirculated slurry is deficient in solid calcium sulfite crystal seeds and the calcium sulfite in the liquor preferentially precipitates on, and blinds, the alkaline limestone particles. This section describes system behavior during limestone blinding, the conditions leading to it, and recommended solutions for eliminating or avoiding limestone blinding.

SYSTEM BEHAVIOR DURING LIMESTONE BLINDING

Limestone blinding in a scrubber system is normally characterized by the following phenomena:

- Severe drop in slurry pH
- Very insensitive pH response to limestone addition at low pH
- Poor limestone utilization
- High sulfite concentration in the liquor

The first indication of limestone blinding is a precipitous drop in the pH of the recirculating slurry for no apparent reason. In order to control system pH, the operator normally begins to increase the limestone feed rate, leading to poor limestone utilization. Limestone utilization as low as 20 to 25 percent has been observed at Shawnee. While the pH response to the limestone feed rate is normally more sensitive at a low pH range of 4.5 to 5.5 (less limestone buffer) than at a high pH range of 5.5 to 6.5 (more limestone buffer), the response is typically sluggish even at low pH when limestone blinding occurs.

CONDITIONS LEADING TO LIMESTONE BLINDING

The necessary conditions for blinding to occur are:

- Slurry solids deficient in calcium sulfite crystal seeds
- High sulfite concentration and/or supersaturation in the slurry liquid

The slurry solids deficient in calcium sulfite crystal seeds (i.e., high gypsum content) can be a result of forced oxidation or high natural oxidation. Experience at Shawnee indicates that limestone blinding does not occur at sulfite oxidation levels in solids below approximately 85 percent under most of the operating conditions.

When the slurry solids contain an insufficient amount of calcium sulfite crystals, the saturated sulfite in the liquor tends to precipitate preferentially on alkaline solid particles such as limestone, because the solubility of calcium sulfite is a strong function of pH and decreases with increasing pH. Thus, even if the bulk liquor is not supersaturated with sulfite, as may be the case with low bulk liquor pH, supersaturation and precipitation could occur in the high pH region in the vicinity of the limestone particles, causing blinding.

High sulfite concentration or supersaturation can be caused by:

- Insufficient oxidation intensity (affecting both scrubber inlet and outlet)
- High SO_2 make-per-pass (affecting scrubber outlet)

The use of additives, such as adipic acid, enhances the SO_2 removal and increases the SO_2 make-per-pass, thus increasing the potential for limestone blinding.

RECOMMENDED SOLUTIONS

Operating Considerations. Limestone blinding by calcium sulfite is the result of calcium sulfite-deficient slurry solids (high gypsum content) and high calcium sulfite supersaturation (or high sulfite concentration) in the liquor. The latter can be caused by insufficient oxidation intensity, high SO_2 make-per-pass, or both.

Therefore, any measures that can reduce these effects will reduce the chance of limestone blinding. Better oxidation can be obtained by:

- Increasing the air stoichiometry
- Increasing the oxidation tank level to provide a longer air bubble residence time
- Increasing the oxidation tank agitation

This would reduce the sulfite saturation and concentration at the scrubber inlet.

The level of sulfite in the scrubber effluent liquor can be reduced by reducing the inlet liquor sulfite as above or by reducing the SO_2 make-per-pass. Lower SO_2 make-per-pass can be obtained by lowering the flue gas throughput, increasing the slurry flow rate, or both.

Design Considerations. If an in-loop forced oxdiation system with a single tank is desired, then provision should be made:

- To provide an adequate oxidation intensity to minimize sulfite saturation at the scrubber inlet
- To reduce SO_2 make-per-pass (outlet sulfite concentration)

A better solution appears to be the use of two tanks in series, which provide several advantages over the single-tank mode listed in Section 7.

Limestone blinding in the long scrubber effluent line (Sections 6 and 7), which acts as a plug-flow reactor, is unique at Shawnee. In full-scale design, the scrubber effluent piping should be as short as possible to minimize the potential for limestone blinding.

Section 10

DEWATERING CHARACTERISTICS OF ADIPIC ACID-ENHANCED LIME/LIMESTONE SLURRIES

Cylinder settling tests and vacuum funnel filtration tests are routinely conducted in the Shawnee Laboratory to monitor the settling and dewatering characteristics of slurry solids.

In the previous reports (References 2 and 3), a comparison of the results of these monitoring tests from July 1978 through October 1979 was presented for lime and limestone slurry with and without adipic acid addition. It was found that adipic acid has an insignificant effect on the quality of solids (settling rate and filterability), except that the settling rate of oxidized limestone slurry may be retarded.

Table 10-1 has been updated to include additional data, obtained from October 1979 through May 1980, for limestone slurry with adipic acid addition both with and without forced oxidation.

The updated data show a higher average initial settling rate of 0.9 cm/min (0.3 to 1.6 cm/min range) for oxidized limestone slurry with adipic acid, compared to the 0.6 cm/min (0.3 to 0.9 cm/min range) previously reported for the same type of slurry (Reference 3). The average initial settling rate for oxidized limestone slurry without adipic acid remains the same at 1.1 cm/min.

The settling rate of unoxidized limestone slurry again shows essentially no effect from adipic acid. The average settling rate is 0.2 cm/min with or without adipic acid.

Table 10-1

COMPARISON OF SHAWNEE WASTE SLURRY DEWATERING CHARACTERISTICS WITH AND WITHOUT ADIPIC ACID ADDITION

Alkali	Fly Ash Loading(1)	Forced Oxidation	Adipic Acid(2)	Initial Solids Conc., wt %	Initial Settling Rate, cm/min.		Ultimate Settled Solids, wt %		Funnel Test Cake Solids, wt %	
					Avg.	Range	Avg.	Range	Avg.	Range
Limestone(3)	High	Yes	Yes	15	0.9	0.3-1.6	70	57-83	70	59-77
Limestone	High	Yes	No	15	1.1	0.6-1.4	73	62-86	74	65-88
Limestone(3)	High	No	Yes	15	0.2	0.1-0.4	50	37-69	56	48-73
Limestone	High	No	No	15	0.2	0.1-0.5	54	41-67	57	48-66
Lime	High	Yes	Yes	8	1.7	1.5-1.9	55	50-60	61	54-69
Lime	High	Yes	No	8	-	-	-	-	-	-
Lime	High	No	Yes	8	1.5	1.2-1.9	51	44-63	62	48-73
Lime	High	No	No	8	1.5	0.8-2.2	49	43-57	56	50-62

(1) Slurries with high fly ash loading contain about 40 percent fly ash in solids.
(2) Adipic acid concentration range is 300 to 4500 ppm.
(3) Data have been updated to include test results from October 1979 through May 1980.

Section 11

ECONOMICS OF ADIPIC ACID-ENHANCED LIMESTONE SCRUBBING

The economics of adipic acid-enhanced limestone scrubbing has been projected for forced-oxidation systems designed to achieve an average of 90 percent SO_2 removal from high sulfur flue gas. The results indicate that, for the cases studied, both capital and operating costs are approximately 4 to 6 percent lower for adipic acid-enhanced limestone systems than for a limestone system without additive. The major savings are in the reduced limestone requirement and the associated grinding equipment. Additional 1 to 2 percent savings in operating cost result from the reduced quantity of waste solids that need to be disposed of in the adipic acid-enhanced limestone system.

The operating conditions for four study cases, including a limestone case with MgO additive, were prepared by Bechtel and are presented in Table 11-1. The capital investment and revenue requirement were calculated by the Economics Evaluation Section of TVA's Energy Design and Operations Branch using a TVA/Bechtel Design-Economics Computer Program (Reference 8). The results are listed in Table 11-2. The evaluations are based on a 500-MW scrubbing facility incorporating forced oxidation and operating on flue gas from a boiler burning 4 wt % sulfur coal. The capital investment and revenue requirement in Table 11-2 include the dewatering equipment (thickener and filter) but exclude the waste sludge (filter cake) disposal area.

The cases evaluated are:

Case 1 - A limestone base case without additive operated at relatively high limestone stoichiometry and liquid-to-gas ratio to achieve 90 percent SO_2 removal. It should be noted that long-term reliability with this mode of operation has not been demonstrated at Shawnee.

Case 2 - A limestone case with MgO addition. Oxidation of the scrubber bleed stream was chosen because in-loop oxidation is incompatible with magnesium-enhanced scrubbing. As in Case 1, long-term reliability has not been demonstrated at Shawnee for this mode of operation.

Case 3 - A limestone case with adipic acid addition operated at high pH. Although only 800 ppm of adipic acid is required to obtain 90 percent SO_2 removal, degradation of adipic acid at high pH requires about five times the theoretical adipic acid addition rate.

Case 4 - A limestone case with adipic acid addition operated at low pH. For this case, 2000 ppm adipic acid is required. However, the low pH operation requires only 1.4 times the theoretical adipic acid addition rate and 1.05 limestone stoichiometry.

Table 11-1

CONDITIONS FOR ECONOMIC ANALYSIS OF ADIPIC ACID-ENHANCED LIMESTONE SCRUBBING WITH FORCED OXIDATION

Capacity:	500 MW
Coal:	4 wt% sulfur
Scrubber:	TCA with 3 beds, 4 grids, and 5 inches of static height of spheres per bed
SO_2 Removal Efficiency:	90%
Superficial Gas Velocity:	12.5 ft/sec
Number of Trains:	5, including 1 spare train
Solids Dewatering:	To 80 wt% solids by thickener and rotary drum vacuum filter
Onstream Factor:	5500 hours/year
Effluent Hold Tank Residence Time:	5 minutes
Oxidation Tank Residence Time:	5 minutes
Oxidation Tank Level:	18 ft
Air Sparger Pressure Drop:	5 psi
Oxidation Tank Agitator Hp:	0.002 Brake Hp/gal
Solid Sulfite Oxidation:	99%
Air Stoichiometry:	1.7 lb-atoms O/lb mole SO_2 absorbed
Number of Tanks:	2 (effluent hold tank and oxidation tank)

Case No.	1	2	3	4
Alkali	Limestone	Limestone	Limestone	Limestone
Additive	-	MgO	Adipic Acid	Adipic Acid
Additive Concentration, ppm	-	5500(a)	800	2000
Additive Rate, lb/hr	-	104	83.3(b)	53.6(c)
L/G, gal/mcf	58	50	50	50
Limestone Stoichiometry, moles Ca/mole SO_2 absorbed	1.52	1.20	1.20	1.05
TCA Inlet pH	5.8	5.4	5.6	4.8
Mode of Oxidation(d)	in loop	bleed stream	in loop	in loop

Notes:
- (a) Effective Mg^{++} concentration.
- (b) Five times theoretical consumption.
- (c) 1.4 times theoretical consumption.
- (d) In-loop oxidation with two tanks uses an oxidation tank followed in-series by an effluent hold tank where alkali is added. Bleed stream oxidation uses one effluent hold tank in the scrubber loop and one bleed stream tank where air is injected.

As shown in Table 11-2, both the total capital investment and the first year revenue requirement are the lowest for adipic acid-enhanced limestone scrubbing at low pH (Case 4). The total capital investment is reduced by 4.8 percent and the first year revenue requirement reduced by 5.8 percent for the limestone/adipic acid/low pH case (Case 4) compared with the conventional limestone case (Case 1). The revenue requirement includes 14.7 percent annual capital charge.

Total capital investment and operating cost for adipic acid-enhanced limestone at high pH (Case 3) are higher than those for limestone/adipic acid at low pH (Case 4), but are still lower than those for the conventional limestone (Case 1) or the limestone/MgO case (Case 2). Total capital investment is lower by 3.9 percent and the first year revenue requirement is lower by 4.0 percent for Case 3, compared with Case 1.

Note that the capital investment and revenue requirements shown in Table 11-2 are significantly different from those presented by TVA in an earlier session. This is due to differences in process equipment and operating parameters itemized in Table 11-3. The most significant factors are coal sulfur content, scrubber type, superficial gas velocity, L/G, hold tank residence time, and landfill investment (not included in Bechtel comparison). The operating parameters in Table 11-1 were selected to represent conditions which have been tested at Shawnee. If the basic design parameters for the comparison in Table 11-1 were adjusted to be the same as those used in the earlier TVA comparison, the same relative results (i.e., limestone scrubbing with adipic acid addition is slightly more economical than standard limestone scrubbing) would be obtained.

Table 11-4 illustrates the additional savings that result from adipic acid addition. Because of the lower pH operation, and thus lower limestone consumption, the amount of waste solids produced is lower for limestone/adipic acid cases (Cases 3 and 4) than for a limestone case (Case 1). Assuming a landfill disposal cost of $10/dry ton, including 14.7 percent annual capital charge, the first year revenue requirements for the sludge disposal area are 0.97, 0.83, and 0.77 mills/kWh for Cases 1, 3, and 4, respectively. Thus, the total first year revenue requirement is 9.34 mills/kWh for Case 4 compared with 10.06 mills/kWh for Case 1. This is a reduction of 7.2 percent, compared with 5.8 percent when the sludge disposal cost is not included.

It should be noted that the differences in total capital investments and operating costs amoung these cases are small. Furthermore, the cost figures are not meant to be accurate or representative of actual scenarios. The principal conclusion from these evaluations is that adipic acid addition does not increase costs but decreases them slightly on the same comparison basis.

Table 11-2

RESULTS OF ECONOMIC ANALYSIS OF ADIPIC ACID-ENHANCED LIMESTONE SCRUBBING WITH FORCED OXIDATION

Case No.	Additive	Additive Conc.,ppm	Total Capital Investment(b) $MM(1982)	Total Capital Investment(b) $/kW	First Year Revenue Requirement(b)(c) $MM(1984)	First Year Revenue Requirement(b)(c) Mills/kWh
1	-	-	87.40	174.8	25.01	9.09
2	MgO	5500(a)	85.26	170.5	24.15	8.78
3	Adipic Acid	800	83.97	167.9	24.01	8.73
4	Adipic Acid	2000	83.22	166.4	23.56	8.57

Notes: (a) Effective Mg^{++} concentration.
(b) Does not include waste sludge disposal area.
(c) Includes 14.7% annual capital charge.

Raw Material Costs (1984): Limestone - \$8.5/ton
MgO - \$460/ton
Adipic Acid - \$1200/ton

Table 11-3

LIMESTONE PROCESS COMPARISON

Item	Bechtel (Case 3)	TVA
Type scrubber	TCA	Spray tower
Superficial gas velocity	12.5 ft/sec	10 ft/sec
% sulfur in coal (as fired)	4.0	3.36
Effluent hold tank residence time	5 min	12 min
Oxidation tank residence time	5 min	5 min
Air stoichiometry	1.7	2.5
Landfill investment	Not included	Included
Adipic acid	800 ppm	1000 ppm
L/G	50	80
Limestone stoichiometry	1.2	1.2
Adipic acid consumption ratio(actual/theor.)	5	3

Table 11-4

REVENUE REQUIREMENT IN SLUDGE DISPOSAL AREA

Case No.	Filter Cake, dry tons/hr	First Year Revenue Rquirement, Mills/kWh[a] Total Excluding Sludge Disposal[b]	Sludge Disposal[c]	Total
1	48.7	9.09	0.97	10.06
2	41.6	8.78	0.83	9.61
3	41.6	8.73	0.83	9.56
4	38.3	8.57	0.77	9.34

Notes: (a) Includes 14.7% annual capital charge.
Costs are based on 1984 dollars.
(b) From Table 11-2.
(c) Assumes $10/dry ton, including 14.7% annual capital charge.

Section 12

SUMMARY OF CURRENT WORK

Important test results of adipic acid-enhanced limestone scrubbing (with high fly ash loading) on the venturi/spray tower system from October 1979 through May 1980 are summarized below. The summary of previous work through October 1979 has been presented in Section 2.

- Factorial tests with spray tower only (venturi plug wide open with 125 gpm slurry flow) showed that, within the operating ranges of 15 to 75 gal/mcf spray tower liquid-to-gas ratio, 4.6 to 5.9 scrubber inlet pH, and 600 to 2400 ppm adipic acid concentration, each 0.4 unit drop in the scrubber inlet pH requires a 700 ppm increase in adipic acid concentration to achieve a similar percent SO_2 removal.

- Apparent degradation of adipic acid is quenched at low pH.* Without forced oxidation, essentially no degradation occurs at a scrubber inlet pH below about 5.0. Both forced oxidation and high limestone stoichiometry (due to limestone blinding at low pH conditions) contribute to higher adipic acid degradation.

- Operating a scrubber at low pH and high adipic acid concentration can actually result in lower total adipic acid consumption than operation at a high pH and low concentration for the same SO_2 removal.

- The optimum scrubber inlet pH appears to be 5.0 to 5.1 with respect to adipic acid consumption, limestone utilization, and the sensitivity of SO_2 removal to pH and inlet SO_2 concentration.

- Operation with the venturi alone (slurry flow to the spray tower turned off) without forced oxidation indicated that the SO_2 removal levels off at a maximum value of about 65 percent at 2000 to 3000 ppm inlet SO_2 concentration, with 3500 to 4500 ppm adipic acid, 21 gal/mcf liquid-to-gas ratio, 5.1 venturi inlet pH, and 8.3 inches H_2O venturi pressure drop. This mode of operation could, however, be attractive for low-sulfur coal having less than 1000 ppm inlet SO_2 concentration where only 70 percent SO_2 removal is required.

- In an in-loop forced oxidation system, or in a system with high natural oxidation, blinding of alkaline limestone particles by calcium sulfite could occur because of the deficiency in calcium sulfite seed crystals. Operation with two tanks, with forced oxidation in the first tank and limestone added to the second tank, minimizes the potential for limestone blinding.

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* Recent laboratory test results at the University of Texas at Austin (Reference 9) have shown that the adipic acid degradation decreases in the presence of Mn ion, and also decreases with pH when Mn is present. The IERL-RTP pilot plant test results (Reference 10) have identified Mn and Fe ions as possible inhibitors of adipic acid degradation.

- A long scrubber slurry-filled effluent pipeline (a flow configuration which exists at Shawnee on the venturi/spray tower system due to system constraints) is detrimental in that it could act as an effective plug-flow reactor for calcium sulfite precipitation and increase the potential for limestone blinding. Limestone blinding in this manner cannot be totally eliminated by increasing the oxidation intensity because calcium sulfite precipitates before being oxidized.

- Additional data showed that adipic acid only slightly reduces the settling rate of oxidized limestone slurry, to 0.9 cm/min (vs. 0.6 cm/min previously reported) from 1.1 cm/min for oxidized limestone slurry without adipic acid.

- Economic analyses for a TCA system with 90 percent SO_2 removal from 4 percent sulfur coal show that both capital and operating costs, excluding the waste solids disposal area, are approximately 4 to 6 percent lower for limestone scrubbing systems with 800 to 2000 ppm adipic acid than for a limestone system without additive. Additional savings for limestone systems with adipic acid can be realized in the waste solids disposal area because of lower solids production rate.

Section 13

FUTURE SHAWNEE TEST PLAN

The test program for the Shawnee Test Facility, as presently conceived for the remainder of 1980 and 1981, is presented below. The major effort will still be placed on the adipic acid-enhanced limestone scrubbing.

In late-May and the first-half of June 1980, Train 100 was converted from a venturi followed by a spray tower to a spray tower-only system. In addition, the spray tower piping and the internal headers were modified in August 1980 to increase the maximum slurry flow rate from 1600 gpm to 2400 gpm. The following test activities with the spray tower only are either in progress, planned, or suggested:

- Factorial tests with limestone slurry with or without forced oxidation, and with or without adipic acid addition, to expand the existing data base and computer models for predicting SO_2 removal.
- Long-term (500 hours) demonstration tests with the spray tower only using limestone/adipic acid slurry with and without forced oxidation.
- Tests to develop design criteria for the spray tower internals.
- Tests with packings having low pressure drop, high efficiency, and low plugging and scaling potential, such as Glitsch Grid packing.
- Tests with other organic acid additives such as dibasic acid, which is a byproduct of adipic acid manufacture consisting primarily of adipic, glutaric, and succinic acids.
- Tests with low SO_2 during the Boiler No. 10 baghouse acceptance testing.
- Integrated power plant water management testing, such as water reuse and additive recovery.
- Testing with other alkalis, such as water treatment sludge, partially calcined limestone, and hydrated dolomitic lime.

The TCA system (Train 200) was restored from a DOWA basic aluminum sulfate process operating configuration in late-June 1980. The following activities are either proceeding, planned, or suggested:

- Simulation of the two full-scale TCA units operating with adipic acid-enhanced limestone at the Southwest Station of the Springfield City Utilities at Springfield, Missouri, as part of the EPA full-scale adipic acid demonstration program.
- Automatic limestone feed control testing.
- Testing with sodium thiosulfate as an oxidation inhibitor.

- Tests with Glitsch Grid packing in lieu of spheres.
- Tests with other organic acid additives, such as dibasic acid.
- Development of a magnesium or calcium adipate clear liquor scrubbing process.
- Development of other forced oxidation methods.
- Tests with low SO_2 during the Boiler No. 10 baghouse acceptance testing.
- Investigation of the effects of limestone type and grind on SO_2 removal and limestone utilization.

Some of the tests listed above are interchangeable between Train 100 and Train 200.

Section 14

REFERENCES

1. Borgwardt, R. H., "Significant EPA/IERL-RTP Pilot Plant Results," in Proceedings: Industry Briefing on EPA Lime/Limestone Wet Scrubbing Test Programs (August 1978), EPA-600/7-79-092 (NTIS PB 296517), March 1979 (pp.1-9).

2. Head, H.N., et al., "Recent Results from EPA's Lime/Limestone Scrubbing Programs - Adipic Acid as a Scrubber Additive," in Proceedings: Symposium on Flue Gas Desulfurization - Las Vegas, Nevada, March 1979; Volume I, EPA-600/7-79-167a (NTIS PB 80-133168), July 1979 (pp. 342-385).

3. Burbank, D.A., and S.C. Wang, "Test Results on Adipic Acid-Enhanced Lime/Limestone Scrubbing at the EPA Shawnee Test Facility - Second Report," in Proceedings: the Fifth Industry Briefing on IERL-RTP Lime/Limestone Wet Scrubbing Test Programs (December 1979), EPA-600/9-80-032 (NTIS PB 80-199813), July 1980 (pp 27-113).

4. Rochelle, G.T., "The Effect of Additives on Mass Transfer in $CaCO_3$ or CaO Slurry Scrubbing of SO_2 from Waste Gases," *Ind. Eng. Chem. Fundam.*, Vol. 16, No. 1, pp. 67-75, 1977.

5. Rochelle, G.T., "Process Alternatives for Stack Gas Desulfurization by Throwaway Scrubbing," in Proceedings of Second Pacific Chemical Engineering Congress, Vol.I, p. 264, August 1977.

6. Radian Corporation, *Further Study of Adipic Acid Degradation in FGD Scrubbers*, draft final report for EPA Contract 68-02-2608, Task 72, April 18, 1980.

7. Cavanaugh, C.M., *Buffer Additives for Flue Gas Desulfurization Processes*, M.S. Thesis, The University of Texas at Austin, December 1978.

8. Stephenson, C.D., and R.L. Torstrick, "The Shawnee Lime-Limestone Computer Program," in Proceedings: the Fifth Industry Briefing on IERL-RTP Lime/Limestone Wet Scrubbing Test Programs (December 1979), EPA-600/9-80-032 (NTIS PB 80-199813), July 1980 (pp 167-222).

9. Rochelle, G., *Buffer Additives for Stack Gas Desulfurization by CaO/$CaCO_3$ Slurry*, September 1980 Monthly Progress Report, EPA Grant R806743, EPA/IERL, Research Triangle Park, N.C. (R.H. Borgwardt, Project Officer).

10. Borgwardt, R.H., et al., *Limestone Scrubbing of SO_2 at EPA/RTP Pilot Plant*, Progress Report 46, August 1980.

Appendix

CONVERTING UNITS OF MEASURE

Environmental Protection Agency policy is to express all measurements in Agency documents in metric units. In this report, however, to avoid undue cost or lack of clarity, English units are used throughout. Conversion factors from English to metric units are given below:

To Convert From	To	Multiply By
scfm (60°F)	nm^3/hr (0°C)	1.61
cfm	m^3/hr	1.70
°F	°C	(°F-32)/1.8
ft	m	0.305
ft/hr	m/hr	0.305
ft/sec	m/sec	0.305
ft^2	m^2	0.0929
ft^2/tons per day	m^2/metric tons per day	0.102
gal/mcf	$1/m^3$	0.134
gpm	1/min	3.79
gpm/ft^2	1/min/m^2	40.8
gr/scf	g/m^3	2.29
in.	cm	2.54
in. H_2O	mm Hg	1.87
lb	g	454
lb-moles	g-moles	454
lb-moles/hr	g-moles/min	7.56
lb-moles/hr ft^2	g-moles/min/m^2	81.4
lb-moles/min	g-moles/sec	7.56
psi	kPa	6.895

COCURRENT SCRUBBER TESTS

SHAWNEE TEST FACILITY

By

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ABSTRACT

Prototype cocurrent limestone scrubber tests were performed at the Shawnee Test Facility. The initial cocurrent prototype tests consistently achieved greater than 90% SO_2 removal while operating with inlet flue gas SO_2 concentrations ranging from 1500 ppm to 3000 ppm. Although the prototype scrubber tower was reliable, total system reliability was not achieved during the initial tests at 27 ft/sec superficial scrubber gas velocity, primarily because of solids deposits in the mist eliminator and the inline, indirect steam reheater. At a 20 ft/sec superficial gas velocity and with low fly ash loading in the inlet flue gas there were no significant solids deposits in the mist eliminator or reheater. Mist eliminator operation was reliable during operation with high fly ash loadings and a 20 ft/sec superficial gas velocity, but the inline reheater continued to plug with slurry solids. During forced-oxidation tests with a single scrubber hold tank and multiple hold tanks, operating conditions were identified which consistently removed greater than 90% of the SO_2 and oxidized greater than 95% of the calcium sulfite in the scrubber slurry to gypsum.

COCURRENT SCRUBBER TESTS

SHAWNEE TEST FACILITY

INTRODUCTION

In 1978 the Hydro-Filter scrubber train at the Shawnee Test Facility was modified to demonstrate the cocurrent scrubber concept. The design of the modification and original test program plan were based upon results from pilot cocurrent scrubber tests conducted at the Tennessee Valley Authority (TVA) Colbert Pilot Plant. The initial equipment modification and the 12-month test program (August 1978-July 1979) were funded by the Electric Power Research Institute (EPRI) and implemented by TVA. A second period of cocurrent tests (August 1979-July 1980) was funded by TVA, the Environmental Protection Agency (EPA), and the Department of Energy (DOE). These tests were conducted to demonstrate reliable operating conditions and limestone cocurrent scrubber operation with forced oxidation.

This paper summarizes the results of the TVA cocurrent scrubber tests. The highlights of the Colbert pilot plant tests and the EPRI prototype cocurrent scrubber tests are presented as background for this discussion.

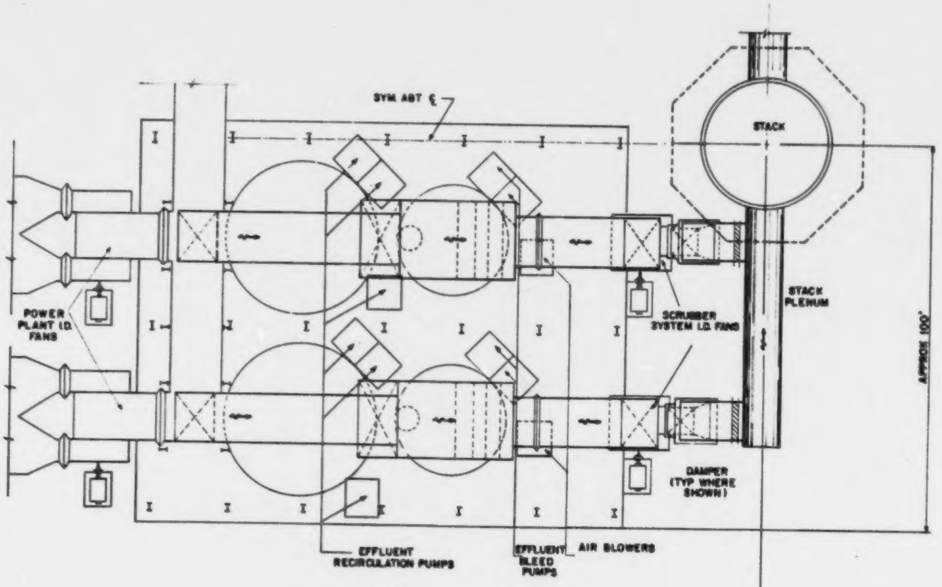
COCURRENT SCRUBBER

Background

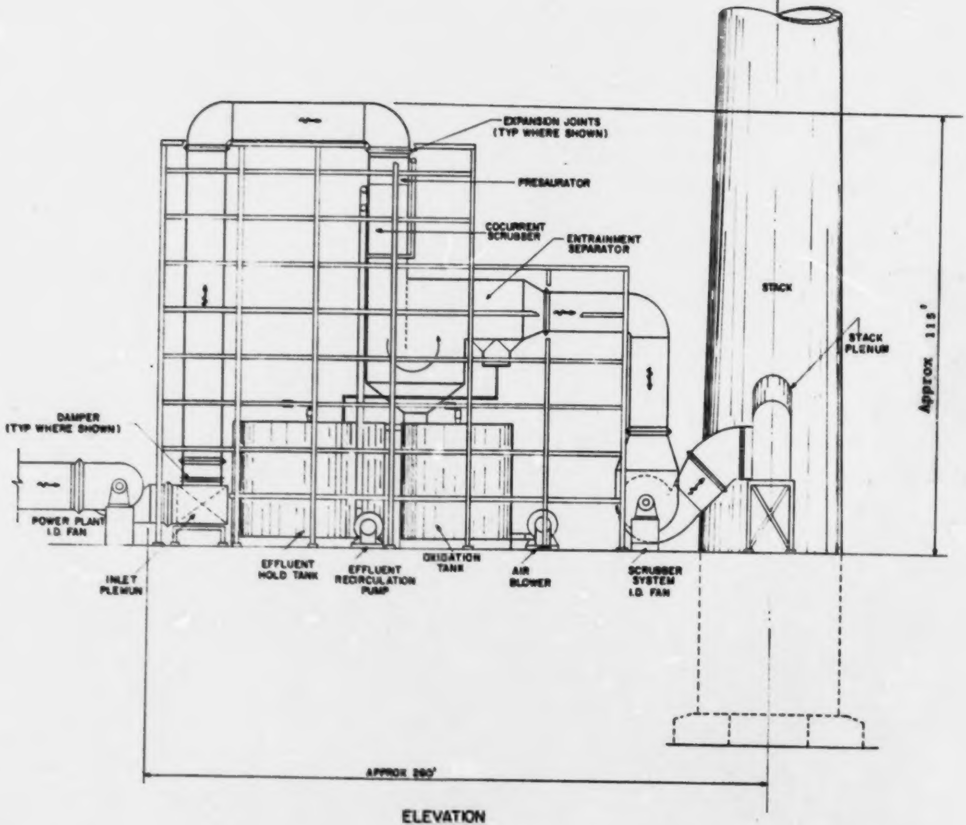
The cocurrent scrubber design as illustrated in Figure 1 has several potential advantages over other commercial FGD scrubber arrangements.

- The equipment configuration is more compatible with most power plant duct and fan arrangements. The gas enters the scrubber at a high elevation and leaves near ground level. The entrainment separator and reheat systems (likely to require the most maintenance) can be near ground level. Likewise, the induced draft (ID) fans can be on the ground and the connecting ductwork to the stack can be shorter and probably less complex.

- The physical arrangement of the cocurrent scrubber causes the gas to change direction in the base of the unit before it enters the mist eliminator. Both the change in direction of the gas and the vertical position of the entrainment separator promote good liquid separation and drainage. Also, a separate mist eliminator wash loop may be used, if needed.


SYM ABT ℄
STACK
POWER PLANT I.D. FANS
STACK PLENUM
SCRUBBER SYSTEM I.D. FANS
APPROX 100'
DAMPER (TYP WHERE SHOWN)
EFFLUENT RECIRCULATION PUMPS
EFFLUENT BLEED PUMPS
AIR BLOWERS

PLAN


EXPANSION JOINTS (TYP WHERE SHOWN)
PRESAURATOR
COCURRENT SCRUBBER
ENTRAINMENT SEPARATOR
STACK
STACK PLENUM
Approx 115'
DAMPER (TYP WHERE SHOWN)
POWER PLANT I.D. FAN
INLET PLENUM
EFFLUENT HOLD TANK
EFFLUENT RECIRCULATION PUMP
OXIDATION TANK
AIR BLOWER
SCRUBBER SYSTEM I.D. FAN
APPROX 280'

ELEVATION

Figure 1. Cocurrent limestone slurry process plan and elevation.

- Scrubbing liquid should coalesce into larger droplets before disengaging from the gas stream near the base of the scrubber and further facilitate efficient operation of the mist eliminator.

- Flooding of the unit with the associated high pressure drop and excessive entrainment of scrubbing slurry (even if grids are added to improve gas-liquid contact) is less likely. Also, during normal cocurrent operation the gas-side pressure loss is lower because some liquid-side energy is recovered.

- Higher gas velocities (small scrubbers) are achieved because of the reduced tendency to flood and because more efficient mist elimination is likely. Therefore, smaller or fewer scrubber modules would be required in a full-scale system.

These potential advantages provided incentive for TVA and EPRI to conduct pilot scrubber studies of the cocurrent scrubber concept with flue gas from a coal-fired boiler at the TVA Colbert pilot plant. Representative results from the Colbert limestone cocurrent scrubber tests are given in Table 1. These results and preliminary economic studies justified prototype testing of the cocurrent scrubber at Shawnee.

TABLE 1. LIMESTONE COCURRENT SCRUBBER TEST RESULTS

COLBERT PILOT PLANT

Inlet SO_2 concentration, ppm	2,461
Outlet SO_2 concentration, ppm	242
Percent SO_2 removal	90
Scrubber superficial gas velocity, ft/sec	28
L/G, gal/kft^3	69
Limestone stoichiometry, mol Ca/mol inlet SO_2	1.26
Height of scrubber, ft	30
Number of grids	5
Depth of each grid, in.	9
Scrubber pressure drop, in. H_2O	15.4

A flow diagram of the Shawnee cocurrent scrubber train as installed for the EPRI cocurrent test program is shown in Figure 2. The scrubber system was designed for operation over a wide range of conditions, which are summarized in Table 2. Figure 3 is a schematic of the Shawnee cocurrent scrubber arrangement.

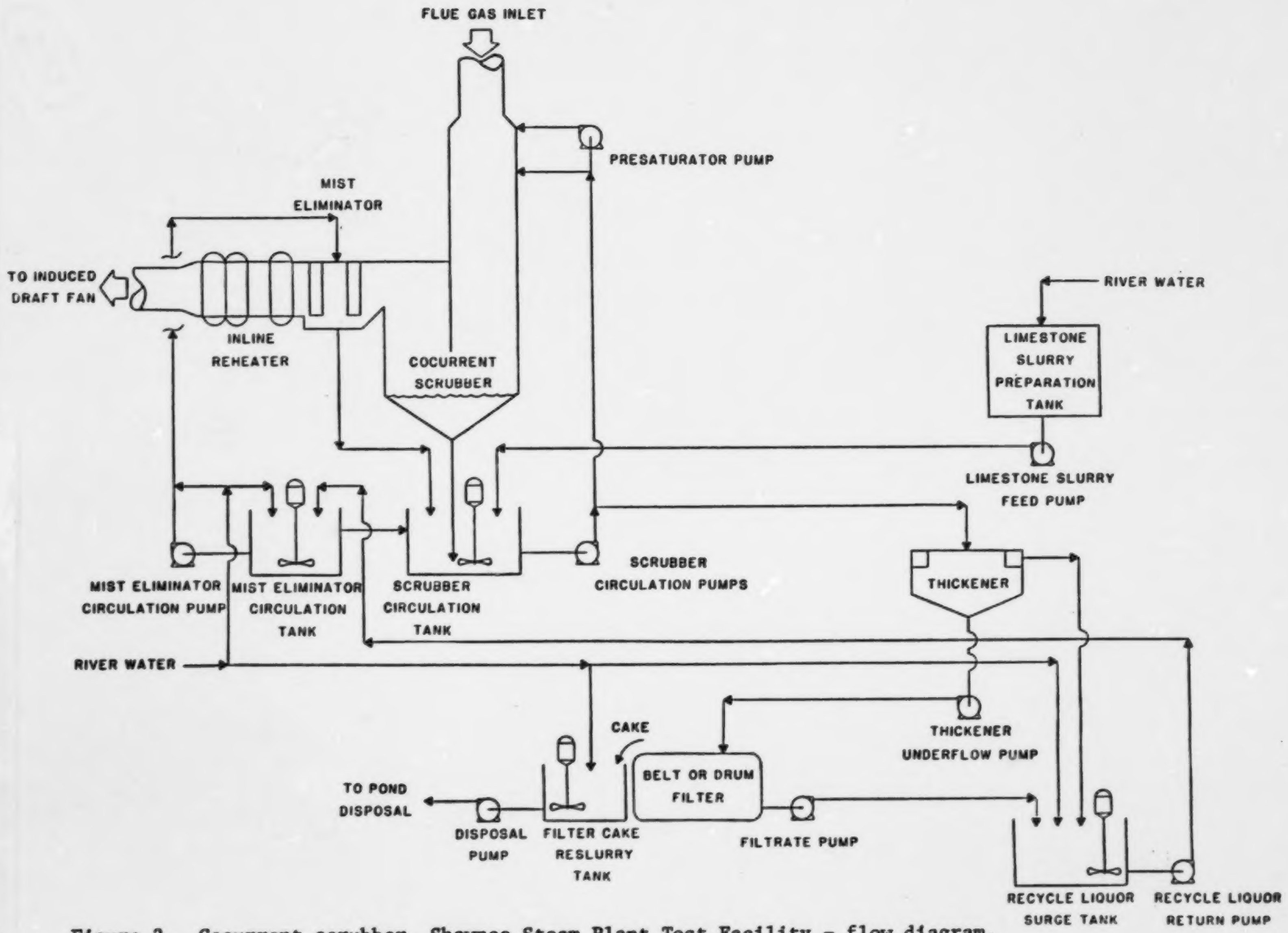

FLUE GAS INLET
PRESATURATOR PUMP
MIST ELIMINATOR
TO INDUCED DRAFT FAN
INLINE REHEATER
COCURRENT SCRUBBER
RIVER WATER
LIMESTONE SLURRY PREPARATION TANK
LIMESTONE SLURRY FEED PUMP
MIST ELIMINATOR CIRCULATION PUMP
MIST ELIMINATOR CIRCULATION TANK
SCRUBBER CIRCULATION TANK
SCRUBBER CIRCULATION PUMPS
THICKENER
RIVER WATER
CAKE
THICKENER UNDERFLOW PUMP
BELT OR DRUM FILTER
TO POND DISPOSAL
DISPOSAL PUMP
FILTER CAKE RESLURRY TANK
FILTRATE PUMP
RECYCLE LIQUOR SURGE TANK
RECYCLE LIQUOR RETURN PUMP

Figure 2. Cocurrent scrubber, Shawnee Steam Plant Test Facility - flow diagram.

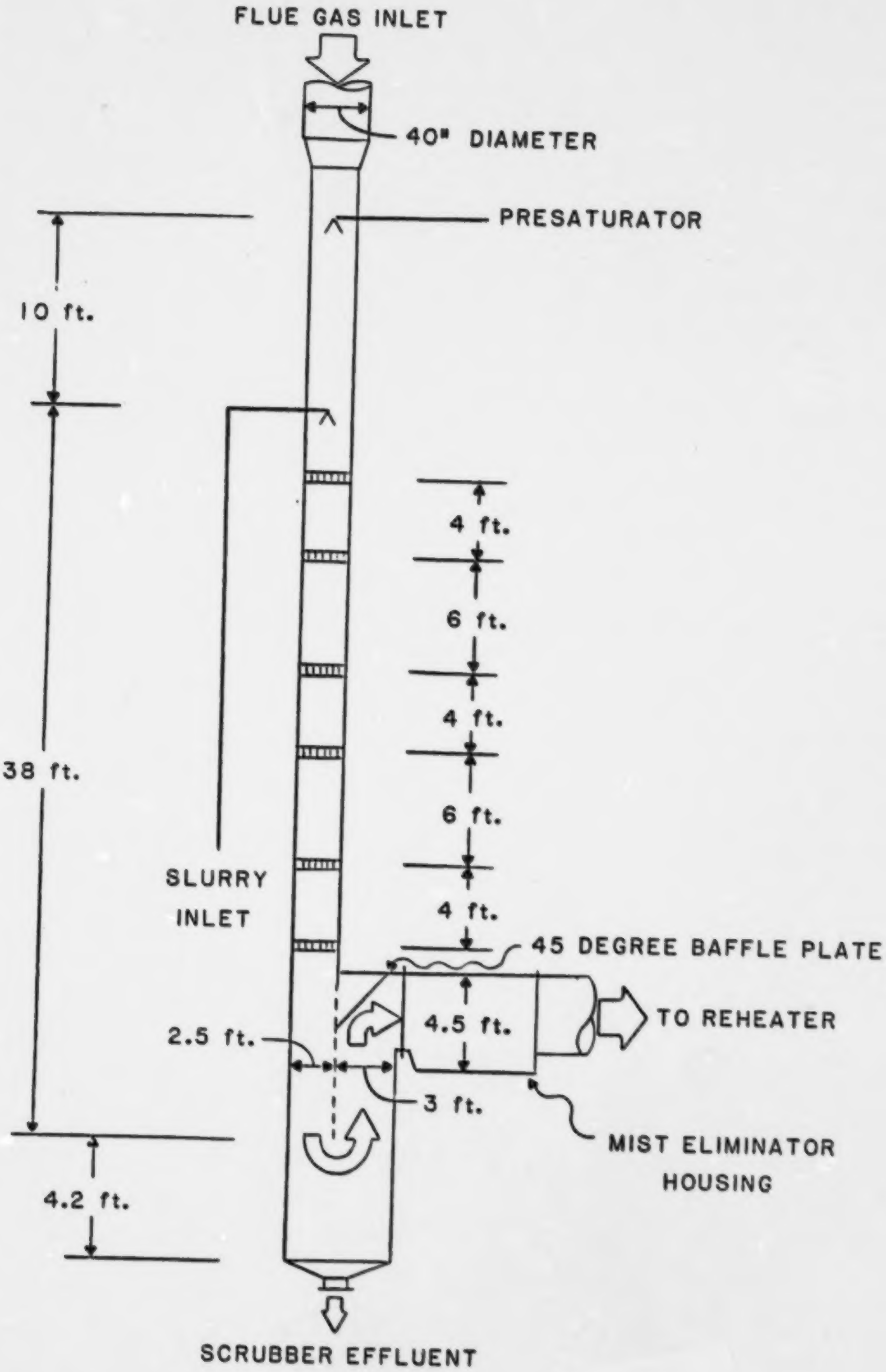

FLUE GAS INLET
40" DIAMETER
PRESATURATOR
10 ft.
38 ft.
4.2 ft.
4 ft.
6 ft.
4 ft.
6 ft.
4 ft.
SLURRY INLET
45 DEGREE BAFFLE PLATE
4.5 ft.
TO REHEATER
2.5 ft.
3 ft.
MIST ELIMINATOR HOUSING
SCRUBBER EFFLUENT

Figure 3. Cocurrent scrubber, schematic.

TABLE 2. SHAWNEE PROTOTYPE COCURRENT SCRUBBER

MAJOR DESIGN PARAMETERS

Design parameter	Range
Scrubber superficial gas velocity, ft/sec	18-31
L/G (at 32 ft/sec gas velocity), gal/kft^3	12-100
Scrubber height, ft	25-45
Number of spray headers	1-4
Number of spray nozzles/header	4-8
Scrubber circulation tank retention time at maximum recirculation rate	6-17

Extensive testing with sodium carbonate, lime, and limestone absorbents was performed during the EPRI-funded program. Detailed results of this program were presented at the EPA Fifth Industry Briefing Conference on Lime/Limestone Wet Scrubbing. Representative results with each of these absorbents are shown in Table 3.

Highlights of the lime/limestone tests included in the EPRI test program follow:

- The gas/liquor contact efficiency of the cocurrent open spray tower (no grids) was inadequate for SO_2 removal greater than 85%.

- Installation of grids in the tower provided effective gas and liquor contact, which increased the SO_2 removal efficiency to greater than 90%.

- Slurry distribution through a single spray header at the top of the scrubber provided higher SO_2 removal than slurry distribution throughout the tower with multiple spray headers.

- Scrubber operating conditions that strongly affected SO_2 removal were gas residence time, recirculated slurry rate, and absorbent stoichiometry. Gas residence time had the strongest effect. For example, at 27 ft/sec gas velocity and 1.0 mol Ca/mol inlet SO_2, the recirculated slurry rates required to maintain 85% SO_2 removal with a 25, 35, and 45 foot scrubber were 2370, 1780, and 1175 gpm respectively.

- SO_2 removal by lime absorbent was slightly lower than that achieved in the Colbert pilot lime tests. At similar operating conditions, the Shawnee scrubber achieved 93% SO_2 removal while the Colbert scrubber achieved 96% SO_2 removal.

TABLE 3. SHAWNEE PROTOTYPE COCURRENT SCRUBBER TEST RESULTS

EPRI TEST PROGRAM

Major test conditions	Absorbents		
	Sodium carbonate	Lime	Limestone
Scrubber physical configuration			
Height, ft	25	45	35
Spray header location(s) (ft from scrubber sump)	25, 15	45	35
Flue gas			
Flow rate, aft^3/min at 300°F	25,000	25,000	25,000
Scrubber superficial gas velocity, ft/sec	26.7	26.7	26.7
Slurry recirculation rate, gpm	1,440	1,200	2,400
L/G, gal/kft^3	72	60	120
Open tower or grid tower	Open tower	Grid	Grid
Scrubber pressure drop, inches H_2O	2-3	3	3
Absorbent stoichiometry			
Mols Na/mol SO_2 absorbed	2.24	-	-
Mols Ca/mol inlet SO_2	-	1.1	1.3
Inlet SO_2 concentration, ppm	2,400	2,800	2,400
SO_2 removal efficiency, %	92	93	90

- During a 350-hour limestone scrubbing test, the prototype cocurrent scrubber consistently averaged 90% SO_2 removal efficiency for each successive 24-hour period. Major scrubber operating conditions for this test were 2500 ppm inlet SO_2 concentration, 27 ft/sec superficial gas velocity, L/G equal to 90 gal/kft^3, 8.3 inches H_2O scrubber pressure drop and limestone stoichiometry equal to 1.3 mol Ca/mol inlet SO_2.

- The scrubber tower with grids operated without scaling and plugging of the tower internals; however, a soot blower was required at the tower inlet to remove solids deposits at the wet/dry interface.

- The total scrubber train did not operate reliably at 27 ft/sec scrubber superficial gas velocity because slurry solids deposits plugged the mist eliminator and reheater.

TVA Cocurrent Scrubber Test Program

After completion of the EPRI program, TVA continued limestone cocurrent scrubber tests with emphasis upon improvement of the mist eliminator and reheater reliability and tests with forced oxidation. The TVA tests were conducted from August 1979 to July 1980. EPA and DOE provided funds for the test program after June 1. The test program was separated into two primary test blocks:

- Mist eliminator reliability tests

- Forced-oxidation tests

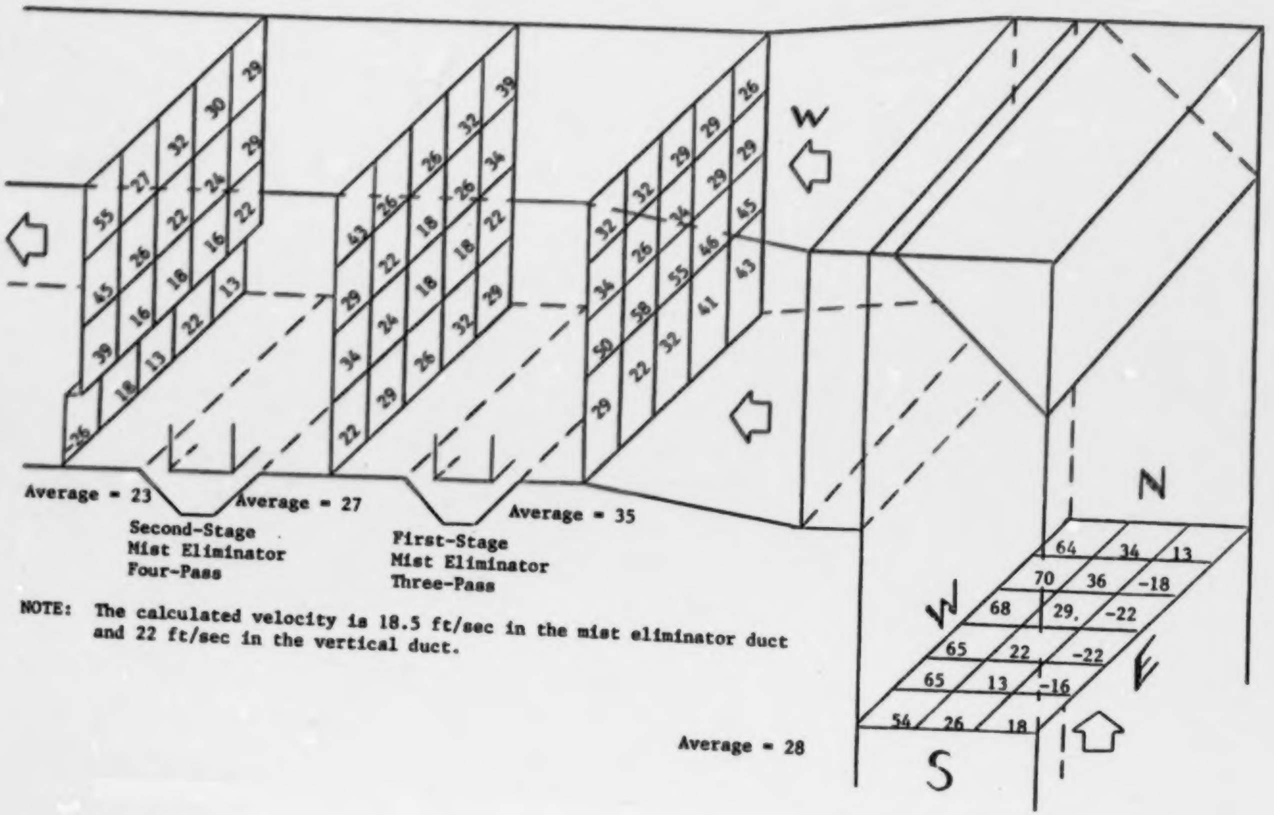
Mist Eliminator Reliability Tests. These tests were performed to determine operating conditions that would provide reliable mist eliminator and reheater operation. Velocity profile determinations upstream and downstream of the mist eliminators indicated that the gas distribution at the mist eliminator entrance was very poor (see Figure 4). The plans for the reliability tests were based primarily upon the hypothesis that improvement of the gas distribution at the outlet of the scrubber and the entrance to the mist eliminators would improve the mist eliminator reliability and efficiency. Improved mist eliminator efficiency would in turn improve the reheater reliability.

Scrubber operating conditions and scrubber equipment were revised during this test block as follows:

- The scrubber superficial gas velocity was lowered from 27 to 20 ft/sec.

- The flue gas source was changed from upstream of the boiler ESP to downstream of the ESP.

- The solids concentration in the recirculated scrubber slurry was reduced from 15% to 10%.


W
N
W
E
S
Average = 23
Second-Stage
Mist Eliminator
Four-Pass
Average = 27
First-Stage
Mist Eliminator
Three-Pass
Average = 35
Average = 28
55 27 32 30 29
45 26 22 24 29
39 16 18 16 22
26 18 13 22 13
43 26 26 32 39
29 22 18 26 34
34 24 18 18 22
22 29 26 32 29
32 32 29 29 26
34 26 34 29 29
50 58 55 46 45
29 22 32 41 43
64 34 13
70 36 -18
68 29. -22
65 22 -22
65 13 -16
54 26 18

NOTE: The calculated velocity is 18.5 ft/sec in the mist eliminator duct and 22 ft/sec in the vertical duct.

Figure 4. Cocurrent scrubber mist eliminator velocity profile (ft/sec). Air only at a scrubber superficial gas velocity of 27 ft/sec.

- Presaturator spray nozzles and a soot blower for solids cleaning were installed at the inlet of the scrubber.

- A 3-pass, open-vane mist eliminator was installed in the outlet duct of the scrubber sump and turning vanes were installed in the 90-degree turn immediately upstream of the mist eliminator.

All of these revisions, except the presaturator installation, were made to decrease the amount of solid and liquid entrainment leaving the scrubber (entering the mist eliminator) and to improve the entrainment removal efficiency of the mist eliminator. Operating conditions and results of this test series are summarized in Table 4. All tests, except test LS-4100C, were performed with low fly ash loading in the flue gas. Tests LS-5000C, 5001C, and 5002C were performed with a 20 ft/sec scrubber gas velocity and tests LS-5010C and 4100C were performed with a 27 ft/sec gas velocity. The presaturator sprays were installed before test LS-5001C. The open-vane mist eliminator and turning vanes were installed before test LS-4100C.

Highlights of these tests are summarized as follows:

- Reduction of the scrubber superficial gas velocity to 20 ft/sec and the fly ash loading to ∿0.1 gr/sft^3 essentially eliminated solids deposits in the mist eliminator and reheater.

- Maximum localized gas velocities in the mist eliminator were reduced from 50-60 ft/sec to 35-40 ft/sec when the scrubber gas velocity was reduced from 27 to 20 ft/sec. (The mist eliminator vendor claims high entrainment removal efficiency at 35 to 40 ft/sec.)

- Solids deposited in the mist eliminator and reheater while operating at 27 ft/sec scrubber gas velocity and low fly ash loadings.

- Solids deposits at the wet/dry interface were controlled by periodically cleaning the area around the presaturator spray nozzles with a soot blower.

- The open-vane mist eliminator and the turning vanes that were installed at the outlet of the scrubber did not improve the mist eliminator and reheater reliability while operating the scrubber at a 27 ft/sec gas velocity.

Further testing at 27 ft/sec scrubber gas velocity was postponed to permit forced-oxidation tests at 20 ft/sec. Future tests at a higher gas velocity (∿30 ft/sec) will be conducted after the scrubber outlet duct is modified to provide better gas distribution in the mist eliminator.

Limestone Cocurrent Scrubber Tests with Forced Oxidation (Single Tank Mode). Limestone scrubbing tests with forced oxidation began in October 1979. The first series of tests was performed with air sparging and limestone

TABLE 4. HIGHLIGHTS OF TVA LIMESTONE MIST ELIMINATOR RELIABILITY TESTS

COCURRENT SCRUBBER

Test Number	LS-5000C	LS-5001C	LS-5002C	LS-5010C	LS-4100C
On-stream time, hr	319	470	348	238	170
Scrubber operating conditions					
Physical configuration					
Height, ft	38	38	38	38	38
Number of grids[a,b]	6	6	6	6	6
Header location[c]	B	B	B	B	B
Pressure drop, in. H_2O	3.5-4.0	3.3-4.3	3.3-4.3	6.6-7.9	7.0-8.5
Flue gas					
Flow rate (inlet), aft^3/min (300°F)	18,750	18,750	18,750	25,000	25,000[d]
Superficial velocity at 125°F, ft/sec	20.0	20.0	20.0	27.0	27.0
Inlet SO_2 concentration, ppm	1,340-2,120	1,320-2,320	1,840-2,440	1,360-2,680	2,000-2,680
Slurry[e,f]					
Recirculation rate, gpm	1,400	1,495	1,495	1,895	1,895
L/G, gal/kft^3	93	100	100	95	95
Laboratory results					
Recirculated slurry					
Solids concentration, wt %	8.3-11.5	8.2-11.5	8.7-11.4	8.7-11.4	14.8-16.2
pH	6.0-6.4	5.93-6.43	5.83-6.40	5.85-6.29	5.84-6.17
Total dissolved solids, ppm	4,100-9,900	2,440-11,295	7,618-17,708	9,423-15,144	5,500-9,545
Solids					
Stoichiometry, mols Ca/mol inlet SO_2	1.12-1.46	1.16-1.52	1.18-1.52	1.16-1.47	1.23-1.44
Thickener underflow					
Solids concentration, wt %	18.8-27.4	16.1-23.0	18.1-27.4	13.6-21.9	24.9-30.0
Filter cake					
Solids concentration, wt %	54.9-77.2	54.7-78.7	56.5-67.3	54.7-62.5	54.3-59.6
SO_2 removal efficiency, %	88-94	92-94	92-94	93-97	92-95

a. Grid elevations: 402, 398, 392, 388, 382, and 378 ft.

b. Depth of grids was 3-3/4 inches/elevation.

c. Header elevation: B, 407 ft.

d. Flue gas with full loading of fly ash. All other tests used flue gas with low loading of fly ash.

e. Includes presaturator slurry.

f. Scrubber recirculation tank slurry depth was 6 ft for all tests except test LS-4100C which was 16.5 ft.

addition in a single scrubber circulation tank as shown in the flow diagram in Figure 5. The scrubber internal arrangement and the range of major operating conditions for this first series are summarized in Table 5.

TABLE 5. MAJOR COCURRENT SCRUBBER OPERATING CONDITIONS

LIMESTONE SCRUBBING TESTS WITH FORCED OXIDATION

SINGLE TANK MODE

Flue gas flow rate (inlet)	18,750 aft^3/min at 300°F
Scrubber gas velocity	20 ft/sec at 125°F
Recirculated slurry rate, gpm	1500
L/G, gal/kft^3	100
Recirculated slurry solids concentration, %	10-15
Scrubber height, ft	38
Limestone stoichiometry, mols Ca/mol inlet SO_2	1.3
Grids	6
Depth of each grid, inches	3-3/4
Air stoichiometry, lb-atoms O/lb-mol SO_2 absorbed	2.0-4.0

The objective of this test block was to define operating conditions that would simultaneously achieve 90% SO_2 removal and oxidize greater than 90% of the calcium sulfite in the recirculated slurry to calcium sulfate dihydrate. Initially screening tests were made to study the effect of oxidation air rate on the scrubber SO_2 removal efficiency and the degree of oxidation. The operating conditions and results of these tests are presented in Table 6. The SO_2 removal efficiency and percent oxidation during several of these tests are briefly summarized below:

Test number LS-	5120C	5121C	5122C	5140C	5130C
Limestone stoichiometry, mols Ca/mol inlet SO_2	1.3	1.3	1.3	1.1	1.1
Oxidation air stoichiometry, lb-atoms O/lb-mol SO_2 absorbed	1.6-2.4	2.6-3.2	2.5	2.3-3.1	2.8-3.7
SO_2 removal efficiency, %	93-97	87-96	88-96	72-85	88-94
Slurry solids oxidation, %	60-81	95-99	66-87	99-100	99-100

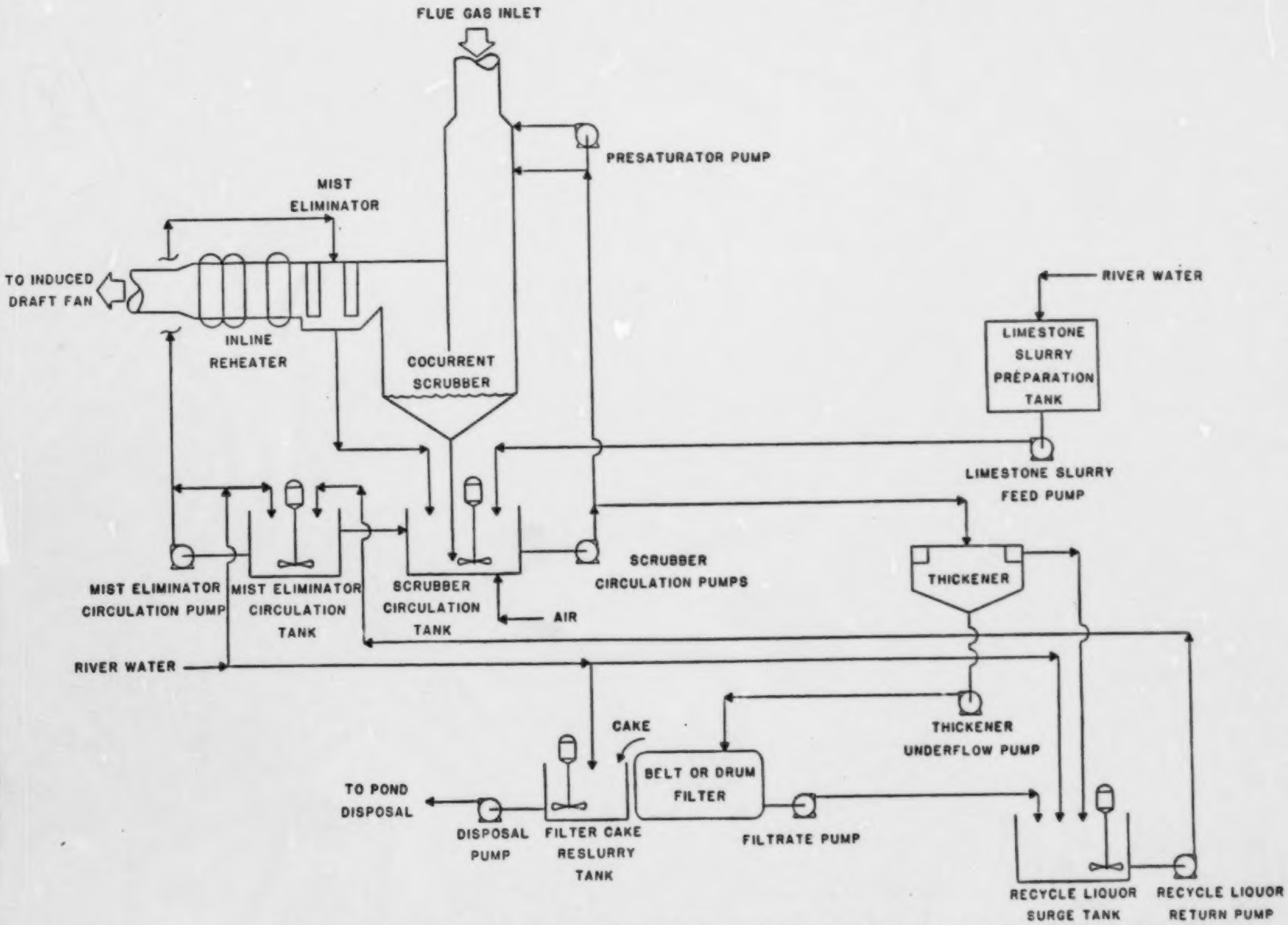

FLUE GAS INLET
PRESATURATOR PUMP
MIST ELIMINATOR
TO INDUCED DRAFT FAN
INLINE REHEATER
COCURRENT SCRUBBER
RIVER WATER
LIMESTONE SLURRY PRÉPARATION TANK
LIMESTONE SLURRY FEED PUMP
SCRUBBER CIRCULATION PUMPS
THICKENER
MIST ELIMINATOR CIRCULATION PUMP
MIST ELIMINATOR CIRCULATION TANK
SCRUBBER CIRCULATION TANK
AIR
RIVER WATER
THICKENER UNDERFLOW PUMP
CAKE
BELT OR DRUM FILTER
TO POND DISPOSAL
DISPOSAL PUMP
FILTER CAKE RESLURRY TANK
FILTRATE PUMP
RECYCLE LIQUOR SURGE TANK
RECYCLE LIQUOR RETURN PUMP

Figure 5. Cocurrent scrubber, Shawnee Steam Plant Test Facility - flow diagram for forced-oxidation tests - single tank mode.

TABLE 6. HIGHLIGHTS OF TVA LIMESTONE COCURRENT SCRUBBER TESTS WITH FORCED OXIDATION - SINGLE TANK MODE

Operating Period	Oct. 24-26	Dec. 13-23	Dec. 27-Jan. 2	Jan. 2-10	Jan. 10-15	Jan. 15-21	Jan. 23-31
Test Number	LS-5100C[a]	LS-5110C[a]	LS-5120C[a]	LS-5121C[a]	LS-5130C[a]	LS-5140C[a]	LS-5122C
Onstream time, hr	47	250	136	192	118	140	178
Scrubber operating conditions							
Physical configuration							
Height, ft	38	38	38	38	38	38	38
Number of stages[b]	6	6	6	6	6	6	6
Number of grids per stage[c]	3	3	3	3	3	3	3
Header location[d]	B	B	B	B	B	B	B
Pressure drop, in. H_2O	3.2-3.6	3.4-4.3	3.5-4.0	3.7-4.3	3.8-4.5	3.6-4.3	3.3-4.0
Flue gas							
Flow rate (inlet), aft^3/min at 300°F	18,750[e]	18,750	18,750	18,750	18,750	18,750	18,750
Superficial velocity, ft/sec at 125°F	20.0	20.0	20.0	20.0	20.0	20.0	20.0
Inlet SO_2 concentration, ppm	1,360-1,680	2,140-3,300	2,120-3,200	2,260-3,320	1,960-2,800	1,900-2,600	1,820-2,720
Slurry[f]							
Recirculation rate, gpm	1,495	1,495	1,495	1,495	1,695	1,695	1,495
L/G, gal/kaft^3	100	100	100	100	113	113	100
Scrubber circulation tank conditions							
Physical configuration							
Slurry depth, ft	10	16.5	16.5	16.5	16.5	16.5	16.5
Agitator speed, rpm	45	68	68	68	68	68	68
Oxidation air							
Rate, sft^3/min	100	250	150	200	200	150	-
Stoichiometry, lb-atoms O/lb-mol SO_2 absorbed	2.1-3.0	2.8-4.0	1.6-2.4	2.6-3.2	2.8-3.7	2.3-3.1	2.4-2.7
Laboratory results							
Recirculated slurry							
Solids concentration, wt %	4.8-8.0	13.6-17.0	13.1-16.7	13.4-17.0	13.0-17.1	13.5-16.6	14.0-16.1
pH	5.9-6.1	5.6-6.1	5.9-6.4	5.5-6.2	5.6-5.9	5.0-5.5	5.7-6.3
Liquor							
Total dissolved solids, ppm	5,072-10,107	5,724-15,618	5,368-7,253	5,428-8,652	7,242-9,642	7,488-11,278	5,500-8,398
Sulfite concentration, ppm	63-145	16-152	9-163	45-253	7-76	416-1,402	34-136
Oxidation, %[g]	92-96	92-99	90-99	85-98	95-100	49-94	91-98
Solids							
Stoichiometry, mols Ca/mol inlet SO_2	1.20-1.38	1.18-1.50	1.31-1.47	1.18-1.35	0.98-1.12	1.08-1.34	1.11-1.43
Oxidation, %[g]	41-48	98-100	60-81	95-99	99-100	99-100	66-87
Thickener underflow							
Solids concentration, wt %	12.0-13.7	18.4-27.3	17.4-30.0	22.6-39.0	16.3-33.5	19.5-27.3	18.2-31.7
Filter cake							
Solids concentration, wt %	-	54.9-89.0	65.4-80.3	63.9-85.3	78.8-89.7	76.0-92.4	67.8-79.2
SO_2 removal efficiency, %	92-94	90-94	93-97	87-96	88-94	72-85	88-96

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a. Turning vanes and open-vane mist eliminator were installed in the 90-degree elbow upstream of the mist eliminator housing.
b. Grid elevations: 402, 398, 392, 388, 382, and 378 ft.
c. Depth of each grid was 1-1/4 inches.
d. Header elevation: B, 407 ft.
e. Flue gas with low-loading of fly ash. All other tests were with flue gas with full-loading of fly ash.
f. Includes 85-95 gpm for presaturator; remaining slurry distributed through six 3-inch, 60-degree spray angle Bete nozzles (ST128FFCN) located at B-header.
g. Percent of total sulfur present as sulfate.

(continued)

TABLE 6 (continued)

Operating Period	Jan. 31-Feb. 5	Feb. 7-28	Feb. 29-Mar. 11	Mar. 11-13	Mar. 14-21	Mar. 21-26
Test Number	LS-5123C	LS-5124C	LS-5150C	LS-5151C	LS-5160C	LS-5161C
Onstream time, hr	119	449	228	64	118	115
Scrubber operating conditions						
Physical configuration						
Height, ft	38	38	38	38	38	38
Number of stages[a]	6	6	6	6	6	6
Number of grids per stage[b]	3	3	6	6	6	6
Header location[c]	B	B	B	B	B	B
Pressure drop, in. H_2O	3.4-3.9	3.5-4.6	6.9-8.2	8.8-9.4	8.3-9.1	8.4-9.1
Flue gas[d]						
Flow rate (inlet), aft^3/min at 300°F	18,750	18,750	18,750	18,750	18,750	18,750
Superficial velocity, ft/sec at 125°F	20.0	20.0	20.0	20.0	20.0	20.0
Inlet SO_2 concentration, ppm	2,200-2,560	2,100-3,000	2,140-3,120	1,900-2,480	1,880-2,540	2,160-2,600
Slurry[e]						
Recirculation rate, gpm	1,495	1,495	1,495	1,885	1,885	1,885
L/G, gal/kaft^3	100	100	100	126	126	126
Scrubber circulation tank conditions						
Physical configuration						
Slurry depth, ft	16.5	16.5	16.5	16.5	16.5	16.5
Agitator speed, rpm	68	68	68	68	68	68
Oxidation air						
Rate, sft^3/min	-	-	-	-	-	-
Stoichiometry, lb-atoms O/lb-mol SO_2 absorbed	2.8-3.1	3.2-3.6	2.3-2.7	2.4-2.7	3.3-3.9	2.9-3.2
Laboratory results						
Recirculated slurry						
Solids concentration, wt %	14.3-16.7	12.3-17.1	13.1-16.7	13.9-16.4	13.9-17.4	13.9-16.3
pH	6.0-6.2	5.5-6.1	5.0-6.1	5.0-5.3	5.1-5.7	5.4-5.8
Liquor						
Total dissolved solids, ppm	5,714-10,531	4,235-9,935	4,523-10,401	9,524-11,144	8,800-12,890	8,725-12,852
Sulfite concentration, ppm	36-100	18-470	145-1,031	14-510	7-118	18-45
Oxidation, %[f]	93-97	80-99	69-92	75-99	90-99	96-99
Solids						
Stoichiometry, mols Ca/mol inlet SO_2	1.30-1.38	1.18-1.51	0.98-1.32	0.98-1.18	0.99-1.21	1.03-1.26
Oxidation, %[f]	82-92	92-99	85-96	99	98-100	99-100
Thickener underflow						
Solids concentration, wt %	20.8-39.4	19.2-31.1	15.6-39.4	23.2-33.7	16.4-31.0	22.3-34.2
Filter cake						
Solids concentration, wt %	53.7-78.6	73.7-87.7	70.1-87.5	78.4-80.7	85.8-89.0	78.1-90.6
SO_2 removal efficiency, %	91-95	87-93	73-90	89-93	90-95	94-96

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a. Grid elevations: 402, 398, 392, 388, 382, and 378 ft.
b. Depth of each grid was 1-1/4 inches.
c. Header elevation: B, 407 ft.
d. Flue gas with full-loading of fly ash.
e. Includes 85-95 gpm for presaturator; remaining slurry distributed through six 3-inch, 60-degree spray angle Bete nozzles (ST128FFCN) located at B-header.
f. Percent of total sulfur present as sulfate.

These tests demonstrated the degree of difficulty associated with simultaneously achieving greater than 90% SO_2 removal and greater than 90% oxidation. In several of the tests, particularly LS-5140C, limestone blinding apparently occurred and the SO_2 removal efficiency of the scrubber decreased. This phenomenon has been explained by the hypothesis that high liquor sulfite concentration (1400 ppm SO_3 in test LS-5140C) and low solid-phase sulfite concentration combine to promote precipitation of calcium sulfite on the surface of the limestone. If this occurs, the limestone dissolution rate, the overall rate-controlling mechanism of this process, decreases and, consequently, the SO_2 removal efficiency drops.

Laboratory analyses (scanning electron microscope examination) clearly indicate that the limestone in the slurry from these tests is not physically blinded. Limestone addition to the scrubber circulation tank during these tests did not, however, provide the expected increase in slurry pH and SO_2 removal efficiency. Further laboratory investigation is needed in an attempt to fully explain this process problem.

The next series of tests (LS-5155, -5200, -5210, and -5201) were performed with the air stoichiometry controlled at 3.0 lb-atoms O/lb-mol SO_2 absorbed. The limestone stoichiometry was varied from 1.1 to 1.3 mol Ca/mol inlet SO_2. Also, after test LS-5155, the depth of the grids in the scrubber was increased from 3-3/4 to 7-1/2 inches. Operation with thicker grids permitted a decrease in L/G to 85 gal/kft^3 without reducing the SO_2 removal below 90%. The major operating conditions and results of this test series are summarized in Table 7. The performance of the scrubber is briefly summarized below:

Test number LS-	5155	5200	5210	5201
Mode	Single-tank	Single-tank	Single-tank	Single-tank
Inlet SO_2 concentration, ppm	2,060-2,600	1,960-2,600	1,800-2,640	1,560-2,480
SO_2 removal efficiency, %	97	90-93	91-94	90-94
Limestone stoichiometry, mols Ca/mol inlet SO_2	1.11-1.16	1.08-1.18	1.20-1.49	1.06-1.13
Air stoichiometry, lb-atoms O/lb-mol SO_2 absorbed	2.9-3.2	2.9-3.2	2.8-3.2	2.7-3.2
Liquor to gas ratio, gal/kft^3	126	87	85	85
Scrubber pressure drop, inches H_2O	8.4-10.7	6.3-7.0	6.4-7.3	6.2-7.0
Oxidation, %				
Liquor phase	87-98	88-98	90-100	86-100
Solid phase	99-100	98-100	99-100	99-100

All of these tests consistently achieved greater than 90% SO_2 removal and greater than 97% oxidation of the slurry solids. Apparently the higher air stoichiometry prevented limestone blinding during these tests.

TABLE 7. HIGHLIGHTS OF TVA LIMESTONE TESTS WITH FORCED OXIDATION IN A SINGLE TANK MODE

Operating Period	April 30-May 5	May 5-May 9	May 9-May 16	May 16-May 20
Test Number	LS-5155	LS-5200	LS-5210	LS-5201
Onstream time, hr	97	99	167	87
Scrubber operating conditions				
Physical configuration				
Height, ft	38	38	38	38
Number of stages[a]	6	6	6	6
Number of grids per stage[b]	6	6	6	6
Header location[c]	B	B	B	B
Pressure drop, in. H_2O	8.4-10.7	6.3-7.0	6.4-7.3	6.2-7.0
Flue gas[d]				
Flow rate (inlet), aft^3/min at 300°F	18,750	18,750	18,750	18,750
Superficial velocity, ft/sec at 125°F	20.0	20.0	20.0	20.0
Inlet SO_2 concentration, ppm	2,060-2,600	1,960-2,600	1,800-2,640	1,560-2,480
Slurry[e]				
Recirculation rate, gpm	1,885	1,300	1,270	1,265
L/G, gal/$kaft^3$	126	87	85	85
Make-per-pass, milli-g-mol SO_2 absorbed/liter slurry	4.3-5.1	5.5-7.1	4.9-7.3	4.9-7.0
Scrubber circulation tank conditions				
Physical configuration				
Slurry depth, ft	16.5	16.5	16.5	14.5
Agitator speed, rpm	68	68	68	68
Oxidation air				
Rate, sft^3/min	190-210	160-200	150-200	140-190
Stoichiometry, lb-atoms O/lb-mol SO_2 absorbed	2.9-3.2	2.9-3.2	2.8-3.2	2.7-3.2
Laboratory results				
Recirculated slurry				
Solids concentration, wt %	13.8-16.8	12.1-17.3	11.2-16.6	13.3-16.8
pH	5.2-5.8	5.2-5.6	5.4-5.9	5.4-5.8
Liquor				
Total dissolved solids, ppm	8,040-9,953	5,013-13,917	8,530-15,366	10,479-15,922
Sulfite concentration, ppm	45-167	23-158	0-136	0-158
Oxidation, %[f]	87-98	88-98	90-100	86-100
Solids				
Stoichiometry, mols Ca/mol inlet SO_2	1.11-1.16	1.08-1.18	1.20-1.49	1.06-1.13
Limestone utilization, %	-	76.9-83.3	62.5-76.9	71.4-85.1
Oxidation, %[f]	99-100	98-100	99-100	99-100
Thickener underflow				
Solids concentration, wt %	25.5-33.7	15.2-37.9	18.7-41.1	19.6-26.3
Sludge cake				
Solids concentration, wt %	81.5-89.3	77.5-89.7	79.3-87.7	76.4-82.0
SO_2 removal efficiency, %	97	90-93	91-94	90-94

a. Grid elevations: 402, 398, 392, 388, 382, and 378 ft.
b. Depth of each grid was 1-1/4 inches.
c. Header elevation: B, 407 ft.
d. Flue gas with full loading of fly ash.
e. Includes 65-100 gpm for presaturator; remaining slurry distributed through six 3-inch, 60-degree spray angle Bete nozzles (ST128FFCN) located at B-header.
f. Percent of total sulfur present as sulfate.

The percent solids in the gypsum filter cake produced during these tests varied from 76% to 90%. A typical composition of the solids produced is summarized below:

	Weight % (dry)
$CaSO_4 \cdot 2H_2O$	59.5
$CaSO_3 \cdot 1/2H_2O$	0.1
$CaCO_3$	13.4
Fly ash	27.0

The settling rate of the solids varied from about 0.1 cm/min at 65% oxidation to 1.0-2.5 cm/min near complete oxidation. Figure 6 is a plot of the solids settling rate versus percent oxidation, which was generated from solid settling tests performed with oxidized slurry from all of the cocurrent scrubber forced-oxidation tests.

Limestone Cocurrent Scrubber Tests with Forced-Oxidation (Multiple Tank Mode). Following the single tank mode tests, the scrubber circulation equipment was modified to permit operation with multiple hold tanks. A flow diagram of the scrubber train in the multiple tank mode is shown in Figure 7. Potential advantages of this mode of operation are:

- A lower pH for the oxidation reaction
- Liquor compositions less likely to promote limestone blinding
- Improved limestone utilization

In this operational mode air is sparged into the first tank, which receives the scrubber effluent. The lower pH of the effluent should provide improved oxidation air utilization because calcium sulfite solubility increases as the pH decreases. Addition of limestone in the second tank after the slurry liquor is oxidized and the liquor sulfite concentration is low should prevent limestone blinding. The multiple tank arrangement partially simulates plug flow and should improve the limestone utilization.

Five forced-oxidation tests that were performed in this test series are summarized in Table 8. Although additional parametric tests should have been performed, an extended period of operation was required for the reliability demonstration test, LS-6150, before the test program was discontinued in July.

In tests LS-6100, -6110, and -6120, the oxidation air stoichiometry was controlled at 3.0, 2.5, and 2.0 lb-atoms O/lb-mol SO_2 absorbed, respectively, while other process control points remained constant, including the limestone stoichiometry at 1.1 mol Ca/mol SO_2 absorbed. (Test LS-6100B was a repeat of

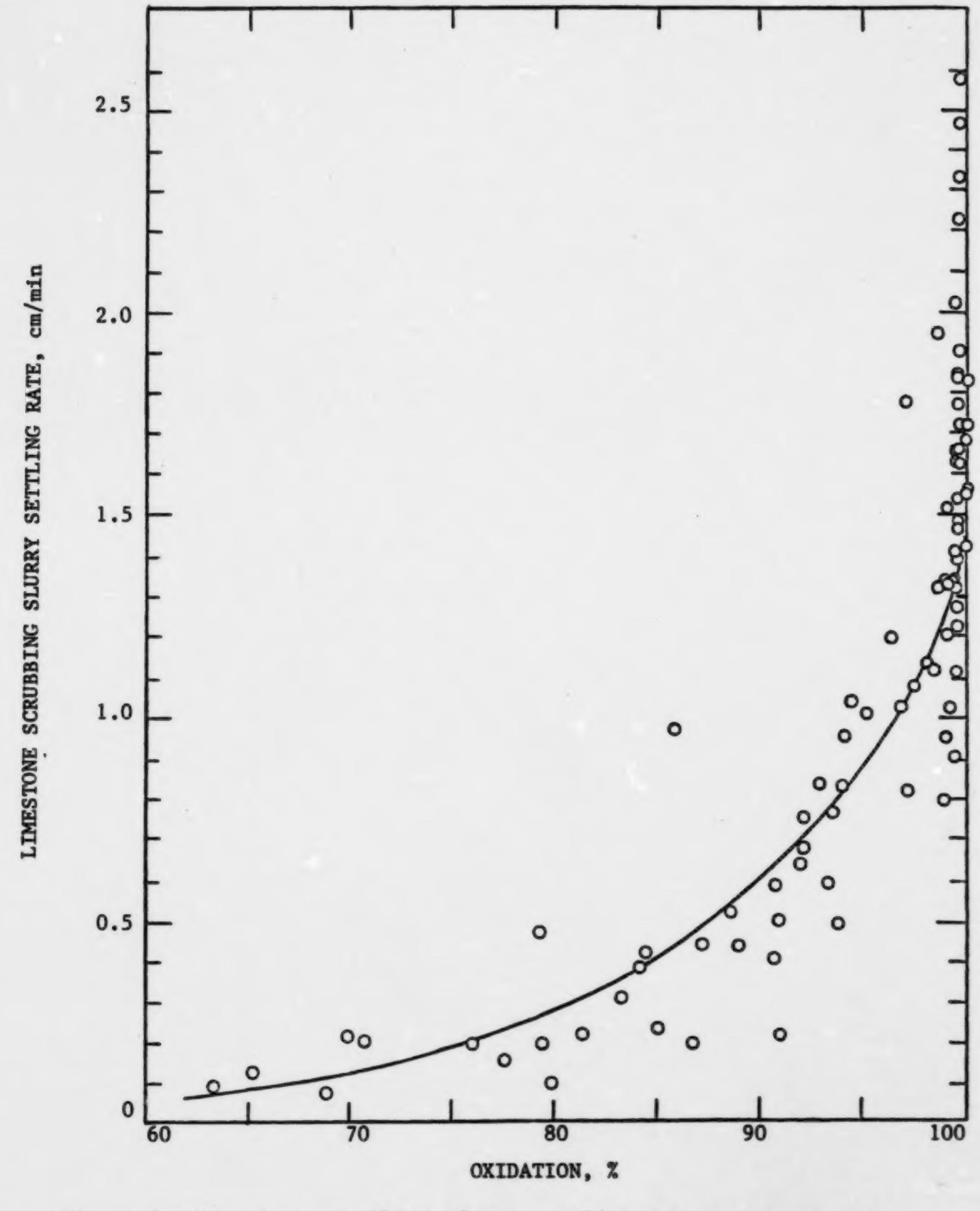

Figure 6. Limestone scrubbing slurry settling rate versus percent oxidation.

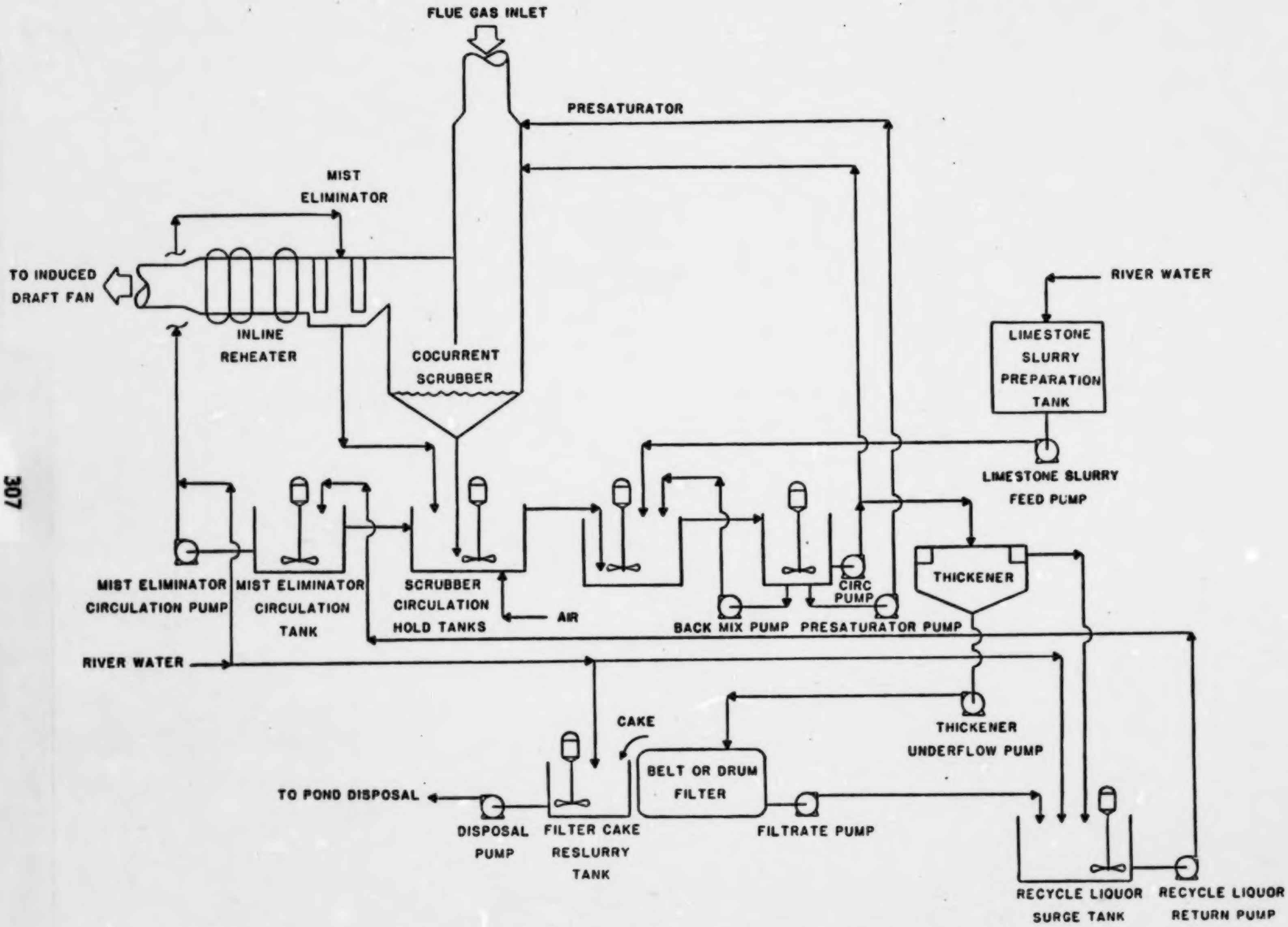
FLUE GAS INLET
PRESATURATOR
MIST ELIMINATOR
TO INDUCED DRAFT FAN
INLINE REHEATER
COCURRENT SCRUBBER
RIVER WATER
LIMESTONE SLURRY PREPARATION TANK
LIMESTONE SLURRY FEED PUMP
MIST ELIMINATOR CIRCULATION PUMP
MIST ELIMINATOR CIRCULATION TANK
SCRUBBER CIRCULATION HOLD TANKS
AIR
BACK MIX PUMP
CIRC PUMP
PRESATURATOR PUMP
THICKENER
RIVER WATER
CAKE
THICKENER UNDERFLOW PUMP
BELT OR DRUM FILTER
TO POND DISPOSAL
DISPOSAL PUMP
FILTER CAKE RESLURRY TANK
FILTRATE PUMP
RECYCLE LIQUOR SURGE TANK
RECYCLE LIQUOR RETURN PUMP

Figure 7. Cocurrent scrubber, Snawnee Test Facility - flow diagram for forced-oxidation tests - multiple hold tanks.

TABLE 8. HIGHLIGHTS OF TVA LIMESTONE COCURRENT SCRUBBER TESTS

WITH FORCED OXIDATION

MULTIPLE TANK MODE

Test number LS-	6100	6110	6120	6100B	6150
On-stream time, hr	143	150	24	40	692
Limestone stoichiometry, mols Ca/mol inlet SO_2	1.0-1.2	1.07-1.10	1.04-1.2	1.07-1.1	1.18-1.42
Air stoichiometry, lb-atoms O/lb-mol SO_2 absorbed	2.8-3.2	2.3-2.7	2.0	2.9-3.1	2.8-3.2
Scrubber L/G, gal/akft^3	85-97	98	98	98	98
Limestone utilization, %		83-91	77-83	83	67-77
Scrubber outlet slurry,					
pH	5.2-5.4	5.3-5.9	5.3	5.3-5.4	5.3-5.6
Liquor					
Sulfite concentration, ppm	160-500	68-588	814	339-452	279-598
Oxidation, %	74-92	71-96	68	79-84	72-87
Solids					
Oxidation, %	99.6-100	99.6-100	99.6	99.5-100	98-100
Recirculated slurry,					
pH	5.8-6.3	5.9-6.3	5.6-6.1	6.0-6.2	5.7-6.4
Solids oxidation, %	98-100	99-100	99.6	99.6-100	94-100
Liquor oxidation, %	88-100	95-100	83-97	98.5-99.4	87-100
Liquor sulfite concentration, ppm	0-181	0-172	45-339	9-68	0-139
SO_2 removal efficiency, %	89-92	89-93	89-91	89-91	92-95

LS-6100.) Operating conditions for the reliability demonstration were selected to ensure that the scrubber SO_2 removal efficiency and percent oxidation were maintained above 90% and 95%, respectively.

Major conclusions and observations from the multiple tank forced-oxidation tests and the reliability demonstration include:

- 90% SO_2 removal efficiency and percent oxidation greater than 98% can be consistently achieved.
- The multiple tank arrangement for these tests does not provide improved limestone utilization. (There appeared to be a slight decrease in limestone utilization, compared with single tank mode with forced oxidation. Additional tests are needed to determine the cause of this unexpected result.)
- Oxidation air utilization is improved in the multiple tank mode. Greater than 95% oxidation was achieved in both the liquor and solid phases of the slurry while operating with 1.1 mol Ca/mol inlet SO_2 and 2.5 lb-atoms O/lb-mol SO2 absorbed. The single tank mode required 3.0 lb-atoms O/lb-mol SO_2 to achieve these conditions.
- Conditions which promote limestone blinding (high SO_3 concentration in the slurry liquor and high percent oxidation of the slurry solids) did not develop until the oxidation air stoichiometry was reduced to 2.0. Limestone blinding occurred with the air stoichiometry controlled at 2.5 during single mode tests.
- The demonstration confirmed the long-term reliability and efficiency of the scrubber tower. The SO_2 removal efficiency was 92% to 95% and the percent oxidation was 98% to 100% during this 700-hour period.
- Although there were no significant solids deposits in the scrubber tower or mist eliminator, the reheater plugged with slurry solids. The deposits in the reheater apparently were caused by higher fly ash loadings in the flue gas (and the resulting higher recirculated slurry density) than were present in the earlier successful reliability demonstration. Additional tests are needed to define operating conditions that do not cause plugging of the inline reheater.

Conclusions. Table 9 is a summary of major design criteria for a cocurrent scrubber system. These criteria apply primarily to the scrubber area and are based upon the results of tests at Shawnee.

TABLE 9. MAJOR DESIGN CRITERIA FOR LIMESTONE COCURRENT SCRUBBER WITH FORCED OXIDATION

Parameter	
Inlet SO_2 concentration, ppm	2,400
Percent SO_2 removal	90
Scrubber superficial gas velocity, ft/sec	20
L/G, gal/kft^3	98
Limestone stoichiometry, mol Ca/mol inlet SO_2	1.3
Number of grids	6
Height of each grid, inches	7.5
Scrubber ΔP, inches H_2O	7.0
Total system ΔP, inches H_2O	13.0
Scrubber height, ft	38
Grid spacing, ft	5
Oxidation tank residence time, min	5
Hold tank residence time, min	10
Air stoichiometry, mols O_2/mol SO_2 removed	1.5
Oxidation efficiency, %	99
Percent solids in throwaway gypsum sludge	80

DOWA PROCESS TESTS

SHAWNEE TEST FACILITY

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ABSTRACT

Dowa dual-alkali process tests at the Shawnee Test Facility were the first application of the Dowa process with flue gas from a coal-fired boiler. The operating conditions were based on operating experience at Dowa facilities at smelter plants, sulfuric acid plants, and oil-fired steam generator plants in Japan.

The initial tests utilized the existing Turbulent Contact Absorber (TCA) in the Shawnee train 200. The maximum SO_2 removal efficiency by the TCA was 85% to 90%. During the TCA testing, problems with gas flow distribution in the absorber were observed. Subsequently, the mobile sphere packing in the TCA was replaced with rigid packing to improve gas flow distribution and gas liquid contact. A factorial absorption test series was conducted using the rigid packing. As a result, operating conditions which will consistently achieve greater than 90% SO_2 removal efficiency were identified.

Performance of the neutralization and gypsum dewatering process steps was generally satisfactory during the absorption tests.

Extensive reliability tests were not conducted; however, no significant reliability problems were identified during the factorial absorption tests.

DOWA PROCESS TESTS

SHAWNEE TEST FACILITY

INTRODUCTION

The Dowa process is a dual-alkali flue gas desulfurization (FGD) process which utilizes basic aluminum sulfate solution for SO_2 absorption and limestone for regeneration of the absorbent. The process was developed by the Dowa Mining Company of Tokyo, Japan, and will be marketed in the United States by the Air Correction Division of UOP, Inc. The process is now in commercial operation in Japan at an oil-fired boiler, smelters, and sulfuric acid plants. The Shawnee prototype Dowa installation is the first test of the Dowa process with flue gas from a coal-fired boiler.

Potential advantages of the Dowa process over the conventional limestone scrubbing process which were justification for the Shawnee tests are:

- Utilizes clear solution scrubbing versus slurry scrubbing to eliminate erosion of equipment and slurry solids buildup on mist eliminator and absorber internals.
- Requires lower limestone stoichiometry.
- Produces a gypsum byproduct which has better dewatering characteristics than unoxidized limestone scrubbing sludge. The Dowa gypsum may be used for wallboard production.

Additional requirements of a Dowa system as compared to a conventional limestone scrubbing system are as follows:

- Includes more equipment and is more complex than a conventional single-loop scrubbing system (excluding sludge mixing and fixation equipment when required as a part of a limestone system).
- Uses an absorbing solution pH of approximately 3 compared to 5 to 6 for a limestone scrubbing system. At the lower pH more acid-resistant materials of construction are required. In areas where carbon steel or a low alloy steel is used in a limestone system, 316L or 317L stainless steel or epoxy resin-lined carbon steel is required.

The Shawnee Dowa process test program was a jointly funded project by the Electric Power Research Institute (EPRI), the Tennessee Valley Authority (TVA), and UOP, Inc. The final month of tests was funded by the Environmental Protection Agency (EPA). The primary purpose of the program was to demonstrate that the process can effectively treat flue gas from a coal-fired boiler. Shawnee train 200, a TCA scrubber system, was modified to the Dowa process configuration and an 8-month test program was conducted. The original program plan included:

1. A 1-month process equipment shakedown and process demonstration at operating conditions recommended by Dowa and UOP.
2. Factorial tests of the absorption process step.
3. Factorial tests of the neutralization and dewatering process steps.

Due to problems with process control and major equipment problems which caused lengthy delays, the neutralization and dewatering tests were eliminated from the test program. However, the neutralization and dewatering sections were operated continuously during all of the tests.

Process Chemistry

The overall chemical reactions in each of the major process steps are:

- Absorption: $Al_2(SO_4)_3 \cdot Al_2O_3 + 3SO_2 \rightarrow Al_2(SO_4)_3 \cdot Al_2(SO_3)_3$ (1)
- Oxidation: $Al_2(SO_4)_3 \cdot Al_2(SO_3)_3 + 3/2O_2 \rightarrow Al_2(SO_4)_3 \cdot Al_2(SO_4)_3$ (2)
- Neutralization: $Al_2(SO_4)_3 \cdot Al_2(SO_4)_3 + 3CaCO_3 + 6H_2O \rightarrow$

  $Al_2(SO_4)_3 \cdot Al_2O_3 + 3(CaSO_4 \cdot 2H_2O) + 3CO_2$ (3)

The process is more accurately defined by the principal intermediate steps within the three process steps, as follows:

- Absorption

$$SO_{2(g)} \rightarrow SO_2\ (\text{diss.}) \qquad (4)$$

$$SO_2(\text{diss.}) \rightleftarrows H_2SO_3(\text{diss.}) \rightleftarrows H^+ + HSO_3^- \qquad (5)$$

$$Al_2(SO_4)_3 \rightleftarrows 2Al^{+++} + 3SO_4^= \qquad (6)$$

In addition to the preceding reactions, the following reactions, which describe the buffering action of basic aluminum sulfate, are important:

$Al^{+3} + x\ OH^- \rightleftarrows Al(OH)_x^{+(3-x)}$ (7)

$Al(OH)_x^{+(3-x)} + H^+ \rightleftarrows H_2O + Al(OH)_{(x-1)}^{+(4-x)}$ (8)

- Oxidation

$HSO_3^- + 1/2O_2 \rightarrow H^+ + SO_4^{-2}$ (9)

- Neutralization

$SO_4^{-2} + CaCO_3(s) + 2H_2O \rightarrow CaSO_4 \cdot 2H_2O\ (s) \downarrow + CO_3^{-2}$ (10)

$CO_3^{-2} + H_2O \rightleftarrows HCO_3^- + OH^-$ (11)

$HCO_3^- + H_2O \rightleftarrows H_2CO_3 + OH^-$ (12)

$H_2CO_3\ (diss.) \rightleftarrows H_2O + CO_2(diss.) \rightleftarrows CO_2(g)$ (13)

The last reaction goes to completion at pH 3.

In summary, sulfur dioxide is absorbed in a solution of basic aluminum sulfate at a pH of approximately 3 [reactions (4) through (8)]. The resultant sulfite in the liquor is oxidized to sulfate by oxygen in the flue gas and in the air which is sparged into the liquor [reaction (9)]. The oxidized liquor is regenerated to basic aluminum sulfate by neutralization with limestone [reactions (10) through (13)]. The gypsum byproduct from the neutralization step is removed by gravitational settling and filtration. The filtrate and clarified liquor are returned to the process.

High SO_2 removal by the process requires the equilibrium of reaction (5) be shifted to the right to allow more HSO_3^- in solution. This is accomplished by more efficient oxidation of the absorber liquor [reaction (9)].

The concentrations of chloride and magnesium in the process liquor are controlled by a purge stream. The aluminum content of the purged liquor is recovered by adding excess limestone to precipitate the aluminum as aluminum hydroxide. The precipitated aluminum is separated from the supernatant liquor and returned to the process. (Equipment for aluminum recovery was not installed at Shawnee.)

Control of the process chemistry requires measurement and control of the total aluminum concentration in the process liquor and the percentage of this aluminum available for the SO_2 absorption reactions. The aluminum concentration is monitored by routine laboratory analysis and controlled by addition of aluminum sulfate solution to the absorber hold tank. The percentage of the aluminum available for absorption is controlled by measurement and control of "% basicity." The concept of % basicity is defined in the following discussion.

Basic aluminum sulfate solution, the absorbent for the Dowa process, is prepared by reacting solutions of aluminum sulfate with limestone to remove sulfate as precipitated gypsum. The limestone is added in less than stoichiometric amounts to prevent converting the aluminum to aluminum hydroxide, which would precipitate. Curve A of Figure 1 shows the pH behavior of an aluminum sulfate solution titrated with either standard acid or standard base. Curve B of Figure 1 shows the behavior of an aluminum sulfate solution that is titrated by incremental additions of powdered calcium carbonate. These results are plotted using the same abscissas as Curve A. The differences between Curve A and Curve B are caused by the presence in the latter case of bicarbonate and carbonate species from the dissolution of the calcium carbonate. These species affect the pH and buffering capacity of the basic aluminum sulfate solution.

The flat portion of the pH curve is the region of interest in the application of basic aluminum sulfate as a scrubbing reagent. For scrubber applications the range of compositions is limited to (N_{OH}/N_{Al}) values of about 0.3 to 1.2, where (N_{OH}/N_{Al}) is the ratio of moles of hydroxide ion per mole of aluminum ion present. The lower limit is chosen to prevent completely exhausting the scrubbing capacity of the liquor, and the upper limit is chosen to prevent potential precipitation of aluminum from the liquor, which would lead to the loss of the aluminum in the gypsum end product produced in the process.

Within the composition range of interest, the liquor pH only changes by 0.2 to 0.5 pH units. This small pH change precludes the use of pH as a process control mechanism. Therefore, in the Dowa process, process control is based upon liquor composition using basicity, B, which is defined as follows:

$$B\ (\%) = \frac{1}{3}\left[\frac{N_{OH}}{N_{Al}}\right]100$$

As examples of the concept of basicity, consider the following:

Compound	Basicity (%)
$Al_2(SO_4)_3$	0
$Al(OH)_x^{+(3-x)}$	$\frac{100x}{3}$
$Al(OH)_3$	100

Three independent means of determining liquor basicity can be used in process control. The liquor basicity is monitored by an in-line basicity analyzer which determines the liquor basicity automatically on a continuous basis. In addition, the liquor basicity can be determined by direct titration in the laboratory or calculated from the results of an analysis of a liquor sample.

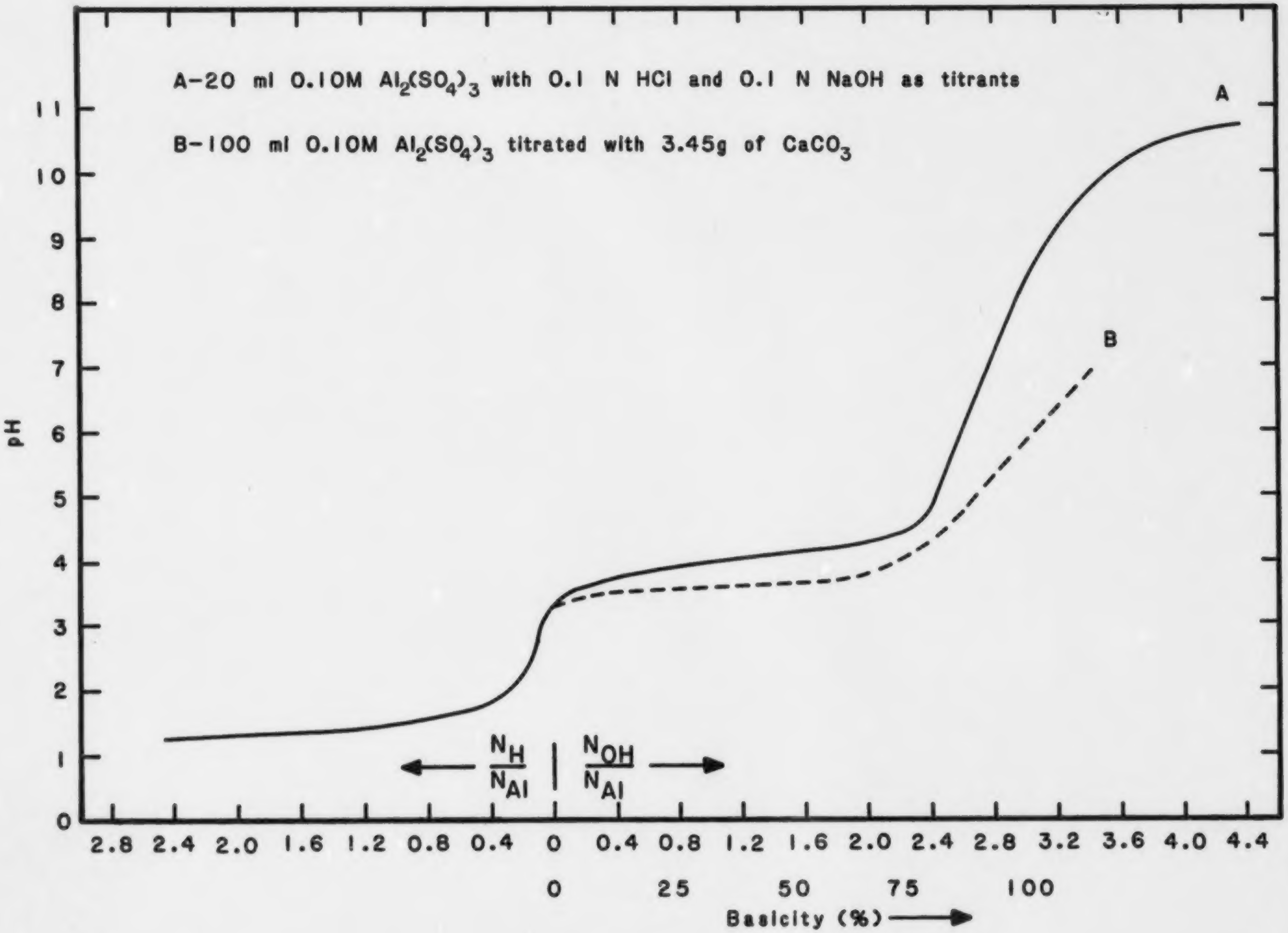

Figure 1. Relationship of pH to basicity and ratios of mols of OH^- and H^+ to mol of Al^{+++} during acid and base titrations of $Al_2(SO_4)_3$ solutions.

SHAWNEE DOWA PROCESS EQUIPMENT DESCRIPTION

Figure 2 is a flow diagram of the Shawnee train 200 after installation of process equipment. The aluminum recovery step was not included in the process demonstration due to limited funds. The major process equipment utilized included:

1. The existing TCA scrubber complete with nitrile foam spheres and a single absorber hold tank (spheres were replaced with rigid packing during the test program)
2. An air sparger system, including a blower and pipe sparger located near the bottom of the absorber hold tank
3. Two neutralization tanks installed in series
4. An existing thickener utilized for initial gypsum dewatering
5. An existing horizontal-belt vacuum filter for final gypsum dewatering
6. A reclaimed absorbent hold tank
7. An aluminum sulfate solution preparation and feed system
8. All process pumps and agitators associated with the above equipment

Sulfur dioxide absorption occurs in the TCA absorber. The oxidation process step occurs in both the absorber and the absorber hold tank. A bleedstream of absorbent is pumped to the neutralizer tanks, where the limestone required for neutralization is added. The neutralizer product overflows from the second neutralizer into a conventional thickener. The thickener overflow is collected in the reclaimed absorbent tank, and the thickener underflow is pumped to the filter for final dewatering of the gypsum byproduct. The filtrate is returned to the reclaimed absorbent tank. A portion of the thickener underflow is recycled to the first neutralization tank to provide gypsum seed crystals for the neutralization/gypsum precipitation step.

The basicity of the absorbent in the absorber loop and the reclaimed absorbent is continuously monitored with an automatic basicity analyzer and routinely analyzed in the test facility laboratory. The basicity of the reclaimed absorbent is controlled by varying the limestone feedrate to the neutralizer tanks. The basicity of the liquor in the absorber loop is controlled by varying the rate of the absorbent purge to the neutralization section.

The aluminum concentration in the absorbent is monitored by laboratory analysis and controlled by aluminum sulfate solution addition to the absorber hold tank.

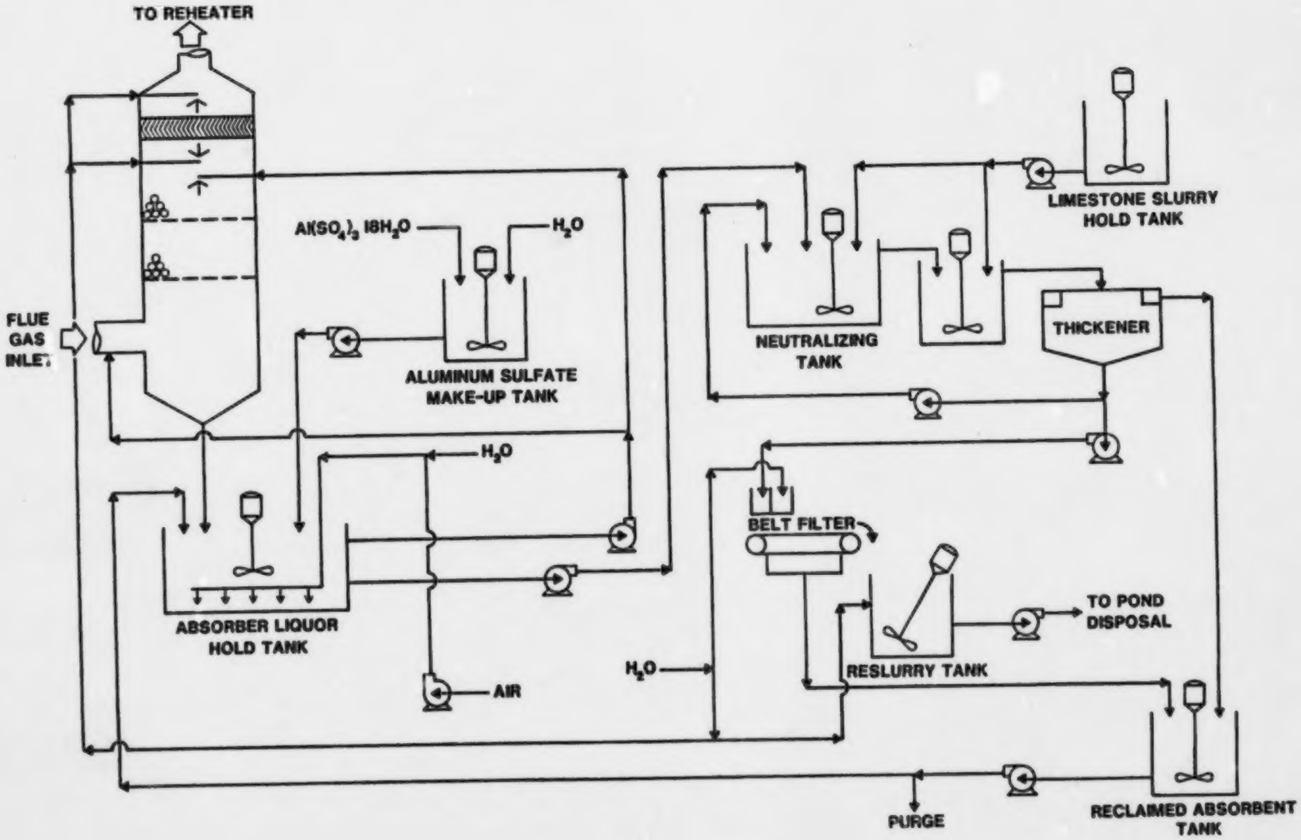
TO REHEATER
FLUE GAS INLET
$Al(SO_4)_3\ 18H_2O$
H_2O
ALUMINUM SULFATE MAKE-UP TANK
H_2O
ABSORBER LIQUOR HOLD TANK
AIR
NEUTRALIZING TANK
LIMESTONE SLURRY HOLD TANK
THICKENER
BELT FILTER
H_2O
RESLURRY TANK
TO POND DISPOSAL
PURGE
RECLAIMED ABSORBENT TANK

Figure 2. Dowa process demonstration flow diagram.

TEST RESULTS

Construction of process equipment changes for the Dowa demonstration was completed in November 1979. Following the completion of construction, numerous equipment-related startup problems plus boiler outages prevented continuous operation of the process demonstration until January 1980.

The major operating conditions selected for the 1-month demonstration are summarized in Table 1. Problems continued to hinder stable continuous operation during the 1-month demonstration.

TABLE 1. DOWA PROCESS DEMONSTRATION

SELECTED ABSORBER OPERATING CONDITIONS

Operating condition	
Inlet flue gas rate, aft^3/min at 300°F	25,000
Inlet flue gas fly ash loading, gr/sft^3	∿0.10
TCA sphere bed static height, inches	
Bed 1	5
Bed 2	5
Bed 3	3
Basicity, %	
Absorber loop	10
Reclaimed absorbent	27
Aluminum concentration, g/ℓ	20
Absorber recycle rate, gal/min	1,250
L/G, gal/kft^3	58
Oxidation air stoichiometry, lb-atoms O/lb-mol SO_2 absorbed	4.0

These problems included:

1. Freezing and plugging the basicity analyzer sample lines caused by poor location of the sample lines and the failure of heat trace material.

2. Unstable standard solutions for calibration of the basicity analyzer caused the process to be controlled either above or below the desired basicity set points. (This problem was not resolved until near the end of the demonstration and may be responsible for scattered test data.)

3. The method for determination of dissolved sulfite in the scrubber liquor was inaccurate. (This problem was solved by addition of iodine to samples to stabilize the sulfite concentration prior to analysis and elimination of the filtration step in the analytical procedure.)

4. Following the initial startup, inspection of the TCA absorber walls and absorber spray nozzles revealed that calcium sulfate scale deposits (from previous limestone scrubbing tests in the TCA) were dissolving in the Dowa liquor, breaking loose from the scrubber internals, and plugging the absorber spray nozzles. Testing was delayed while scale was manually removed from the absorber internals.

Despite the resolution of the above stated problems, SO_2 removal efficiency still did not match the design expectations. In lieu of the continuing process demonstration, a series of TCA screening tests to determine the SO_2 removal efficiency of this absorber over a wide range of flue gas rates, absorbent recirculation rates, and oxidation air flowrates were performed. Attempts to improve control of basicity and oxidation continued during these tests. The SO_2 removal efficiency of these tests continued to be lower than expected from the Dowa process operating with a TCA scrubber in Japan. The TCA static sphere bed depth was increased to 8 inches with little effect on SO_2 removal. The maximum sustained SO_2 removal during this test was 87%. Observation of the sphere action during absorber operation and sphere distribution in the beds after the absorber shutdown indicated that the gas distribution in the TCA was poor. Consequently, poor gas/liquor contact was suspected to be the cause of the low removal efficiency.

The nitrile foam spheres were replaced with fixed-bed packing, and the remainder of the test program was dedicated to absorption studies with this type of packing. Three series of tests were conducted: two series with a 9-foot packing height, and a third with a 6-foot packing height. The ranges of major operating conditions during these tests included:

Flue gas rate, aft^3/min at 300°F	13,000-27,000 (low fly ash loading)
Superficial gas velocity, ft/sec	5.4-11.2
Recirculated absorbent, gal/min	700-1,400
L/G, gal/kft^3	39-126
Absorber ΔP, in. H_2O	1.0-14.7
Absorbent basicity (absorber), %	11.0-35.2

The fixed-bed packing approaches flooding conditions when the absorber is operated at 27,000 aft^3/min (equivalent to 10 MW and an absorber superficial gas velocity of 12 ft/sec). SO_2 removal efficiency did not remain above 90% and steady operation was therefore not possible at a gas rate equivalent to 10 MW.

After the superficial gas velocity was lowered to between 6 and 9 ft/sec, more stable operation and high SO_2 removal efficiency were achieved. For example, 90% to 97% SO_2 removal was achieved while the absorber operating conditions were 20,000 aft^3/min flue gas rate, L/G equal to 55, and pressure drop equal to 9.2 to 10.5 inches H_2O; 93% SO_2 removal was achieved while the absorber operating conditions were 20,000 aft^3/min flue gas rate, L/G equal to 82 gal/kft^3, and pressure drop equal to 9 inches of water; and 93% to 97% SO_2 removal was achieved while the absorber operating conditions were 13,000 aft^3/min flue gas rate, L/G equal to 90 to 125 gal/kft^3, and pressure drop equal to 1.5 to 2.5 inches H_2O.

During the tests with 6 feet and 9 feet of packing in the absorber, an excessive oxidation air stoichiometry was maintained to assure near complete oxidation of the absorber liquor and to enhance SO_2 removal in the absorber. The sulfite concentration in the recirculated absorber liquor during the tests was ∿0 to 60 mg/ℓ. Data collected during the 6-foot fixed-bed packing tests are plotted in Figures 3, 4, and 5. Data collected with 9 feet of packing are being evaluated and will appear in the Dowa project final report.

A typical gypsum byproduct composition during the absorption factorial tests is as follows:

Component	% by wt (dry)
Aluminum	0.3[a]
Calcium	21.8
Carbonate	Nil
Sulfite	∿0.0
Sulfate	53.8
Total solids (wet basis)	81.8
Acid insolubles	Nil

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[a] Gypsum cake washing procedures were not optimized. Lower aluminum concentration is expected with improved cake washing, such as 0.05% Al achieved in commercial facilities in Japan.

The final test was performed with flue gas taken from the duct before the precipitator and thus the gas to the absorber contained full fly ash loading, ∿4.0 gr/aft^3. No significant effect by the fly ash upon the process was observed during this 1-week test.

SUMMARY OF RESULTS

The results of the Dowa test program are summarized as follows:

- Difficulties which were encountered during the Shawnee tests, such as problems with analytical procedures for sulfite and preparation of stable standard solutions for the basicity analyzer calibration, undoubtedly had an adverse effect upon the removal efficiency of SO_2. Also, the apparent poor gas distribution in the Shawnee TCA adversely affected the test results. Although the Dowa process did not effectively remove 90% of the SO_2 in the flue gas from a coal-fired boiler while operating with a mobile-bed scrubber, the quantitative effect of each of these problems upon the observed SO_2 removal efficiency of the TCA is unknown.

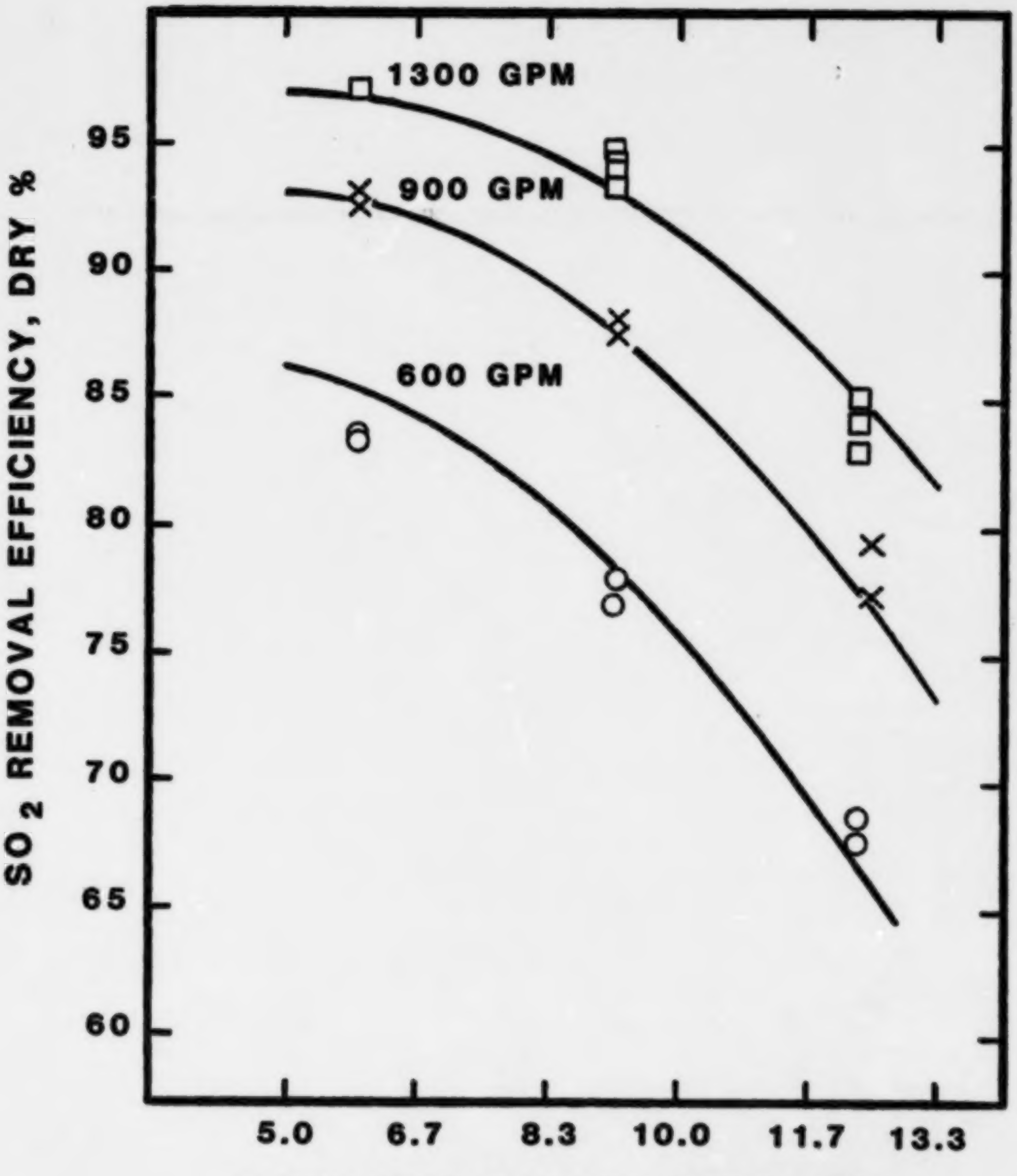

Figure 3. Mass transfer characteristics: Polygrid packing - six foot data.

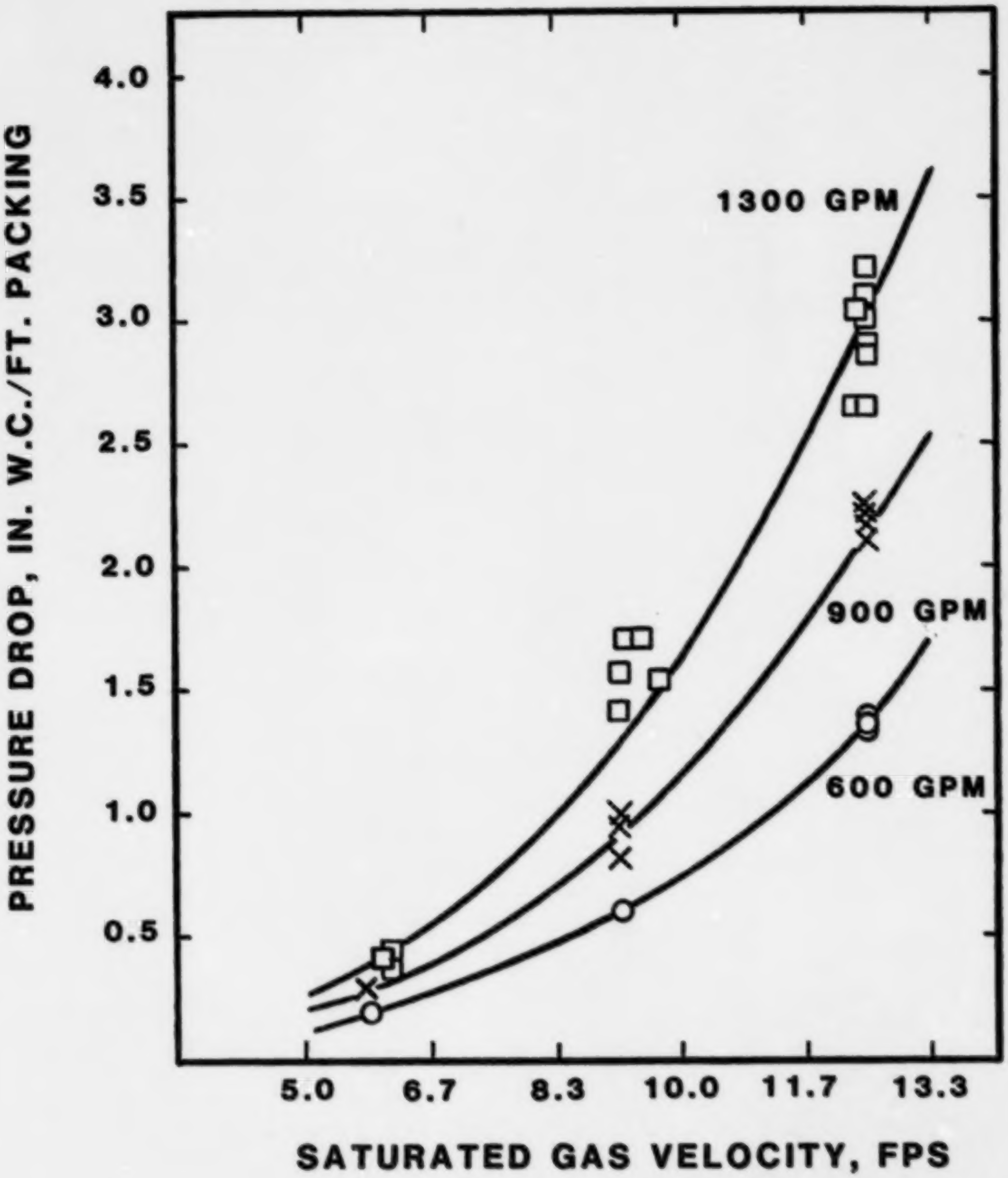

Figure 4. Absorber pressure drop: Polygrid packing - six foot data.

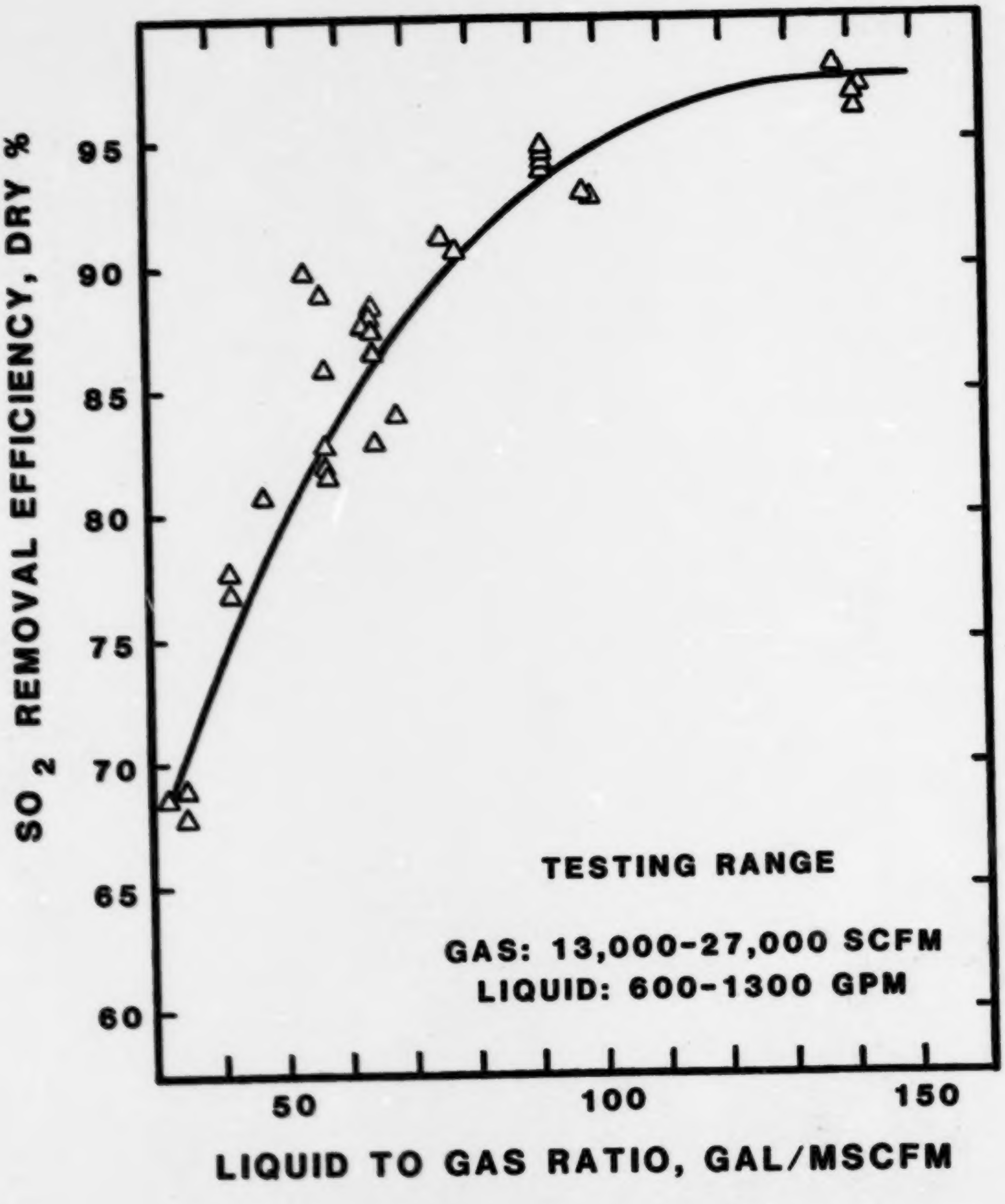

Figure 5. SO_2 removal efficiency versus liquid to gas ratio: Polygrid packing - six foot data.

- The SO_2 removal efficiency of the process is improved to greater than 90% by providing better flue gas and absorbent contact with a fixed-bed packed absorber. For example, with 6 feet of rigid packing in the absorber, 93% SO_2 removal is achieved while operating at 9 ft/sec superficial gas velocity, 82 gal/kft^3 L/G, and 9 inches H_2O scrubber pressure drop.
- The neutralization and dewatering steps of the process can effectively produce a gypsum byproduct.
- High concentrations of fly ash in the process absorbent do not affect the process performance (a preliminary result from a 1-week test).
- There is no scale formation in the absorber.

CURRENT AND FUTURE TESTS

UOP is now performing laboratory studies and installing an integrated pilot plant to further optimize the Dowa process. TVA and UOP have independently proposed further Dowa process tests at Shawnee. These further tests at Shawnee are contingent on completion of these laboratory and pilot tests and an economic evaluation of the process.

F.G.D. EXPERIENCES

SOUTHWEST UNIT 1

N. Dale Hicks, City Utilities, Springfield, Missouri

O. W. Hargrove, Radian Corporation, Austin, Texas

ABSTRACT

City Utilities of Springfield, Missouri, began commercial operation of this F.G.D. system in September, 1977. Two turbulent contact absorber modules are arranged in parallel and utilize a pulverized limestone slurry for SO_2 removal. The scrubbers serve a 195 M.W. unit with a Riley Stoker boiler burning 3.5% sulfur coal. Station design was by Burns & McDonnell, with the Air Correction Division of Universal Oil Products, Inc. responsible for the F.G.D. system on this new facility.

The absorber modules and various support systems have experienced a variety of problems since initial start-up. The more severe problems encountered have been: absorber and demister pluggage; failure of absorber spheres; pipe breakage; control and instrumentation malfunctions; and expansion joint, damper, and duct corrosion. Past and planned efforts to rectify these difficulties, and to improve F.G.D. system reliability, are discussed in detail.

A related problem area has been the continuous monitoring systems for flue gas opacity and SO_2 emissions. Original equipment has proven unsuccessful and the investigation toward a solution, with the aid of a consulting firm, is described.

The station is to be the host facility for an E.P.A. sponsored full scale demonstration of adipic acid as an additive to wet limestone F.G.D. systems. Anticipated results are enchanced efficiency and improved operation of the pollution control facility. Also involved in the project is the Radian Corporation and Universal Oil Products, Inc.

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FGD EXPERIENCES

SOUTHWEST UNIT 1

Southwest Unit 1 is a 194 MW fossil fueled unit owned and operated by City Utilities of Springfield, Missouri. The unit was designed by Burns & McDonnell. Commercial operation of the generating unit began in May, 1976 while the FGD Systems were not fully operational until September 1977.

The steam generator was supplied by Riley Stoker Corporation and was designed to burn high (3.5 to 4.0 percent) sulfur coal. Air Correction Division of Universal Oil Products (UOP) furnished and installed the electrostatic precipitator and the two-module flue-gas scrubber. Each absorber module consists of a presaturator area; three TCA beds, each containing spheres to enhance gas-liquid mixing; two chevron mist eliminator banks; and a reaction hold tank. A common limestone preparation area and sludge dewatering train serve both modules. The design was based on a limestone composition of 98.7 percent $CaCO_3$, 0.7 percent $MgCO_3$, and 0.6 percent inerts.

The attached Figure 1 presents the process flow diagram. Basically, the induced draft (ID) fans pull the flue gas from the boiler through the air heaters and the electrostatic precipitator (ESP) and discharge into the scrubber inlets. Particulates (fly ash) are removed from the flue gas by the ESP and conveyed to dry storage. SO_2 removal is attained in the scrubber with waste products of the reaction removed by continuous outflow from the scrubbers to a thickener tank. The thickener separates the water (supernatant) from the waste solids and recycles the water to the scrubber system. The waste solids are drawn from the thickener and passed across a travelling vacuum belt for further removal of water. This sixty to seventy percent solids waste is then mixed with dry fly ash to produce a fixed material which is landfilled on site.

Limestone is prepared for the scrubbing process by wet grinding in two (2) ball-tube mills. There is no classification of the ballmill outputs so fineness of grind cannot be readily controlled.

PROBLEM AREAS

Numerous problems have plagued the FGD systems at Southwest Unit 1. Some problems have been solved, others are still being dealt with. The following sections will detail the major classifications of problems experienced and findings relative to their evaluations.

Pluggage

Pluggage in both the demister sections and the absorption areas of the scrubbers originally hampered reliable operation of the FGD system. During initial start-up and shakedown of the scrubbers it became readily apparent that the demister wash system was inadequate. The system was designed to operate as a closed loop as shown in Figure 2. Within two weeks of only partial operation the demister chevrons were thickly scaled. Continuous recirculation of the solids-ladened wash water caused a further scrubbing action in the demister area resulting in scale formation and serious plugging.

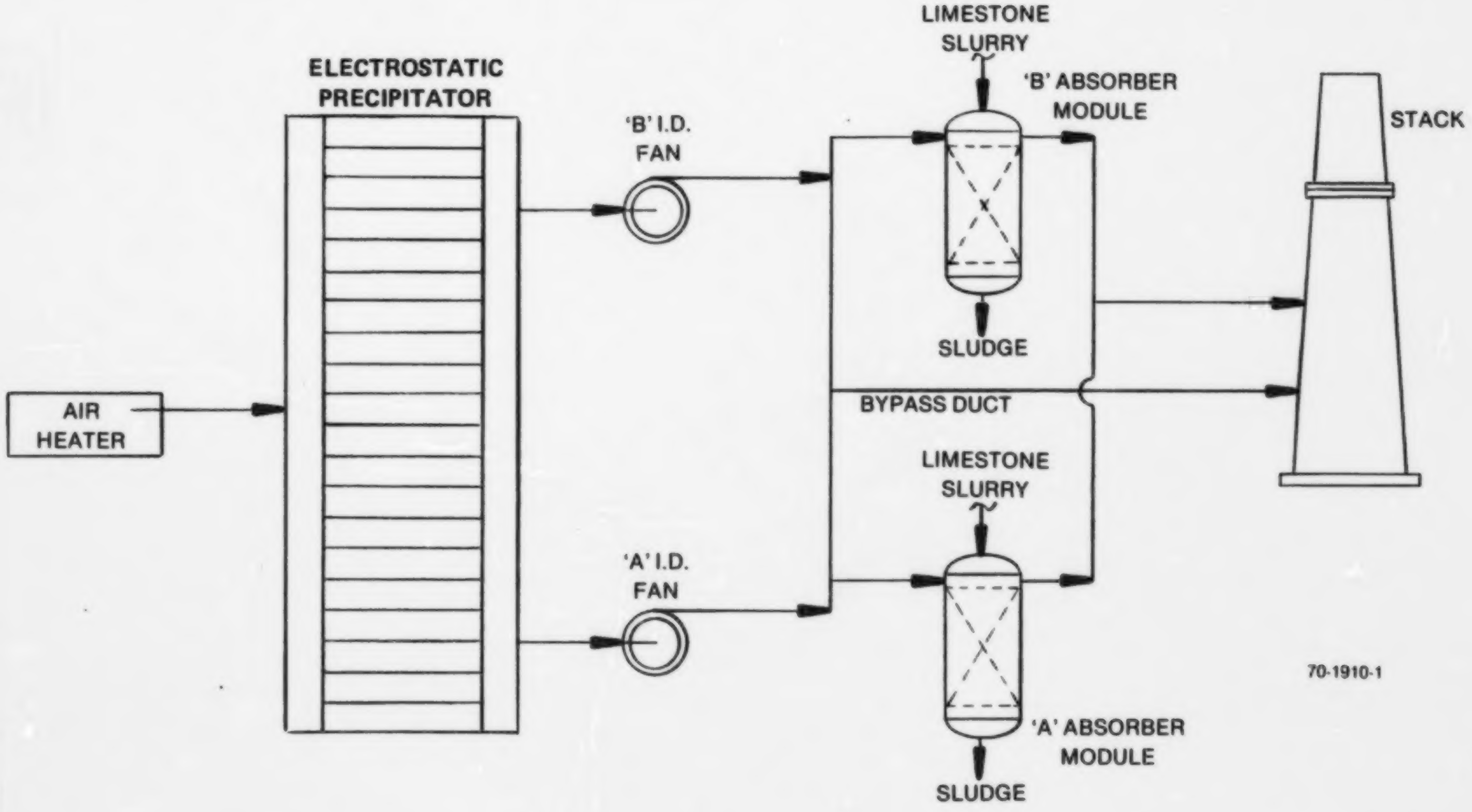

AIR HEATER
ELECTROSTATIC PRECIPITATOR
'B' I.D. FAN
'A' I.D. FAN
LIMESTONE SLURRY
'B' ABSORBER MODULE
SLUDGE
BYPASS DUCT
LIMESTONE SLURRY
'A' ABSORBER MODULE
SLUDGE
STACK
70-1910-1

Figure 1 Southwest Unit 1 Process Flow Diagram.

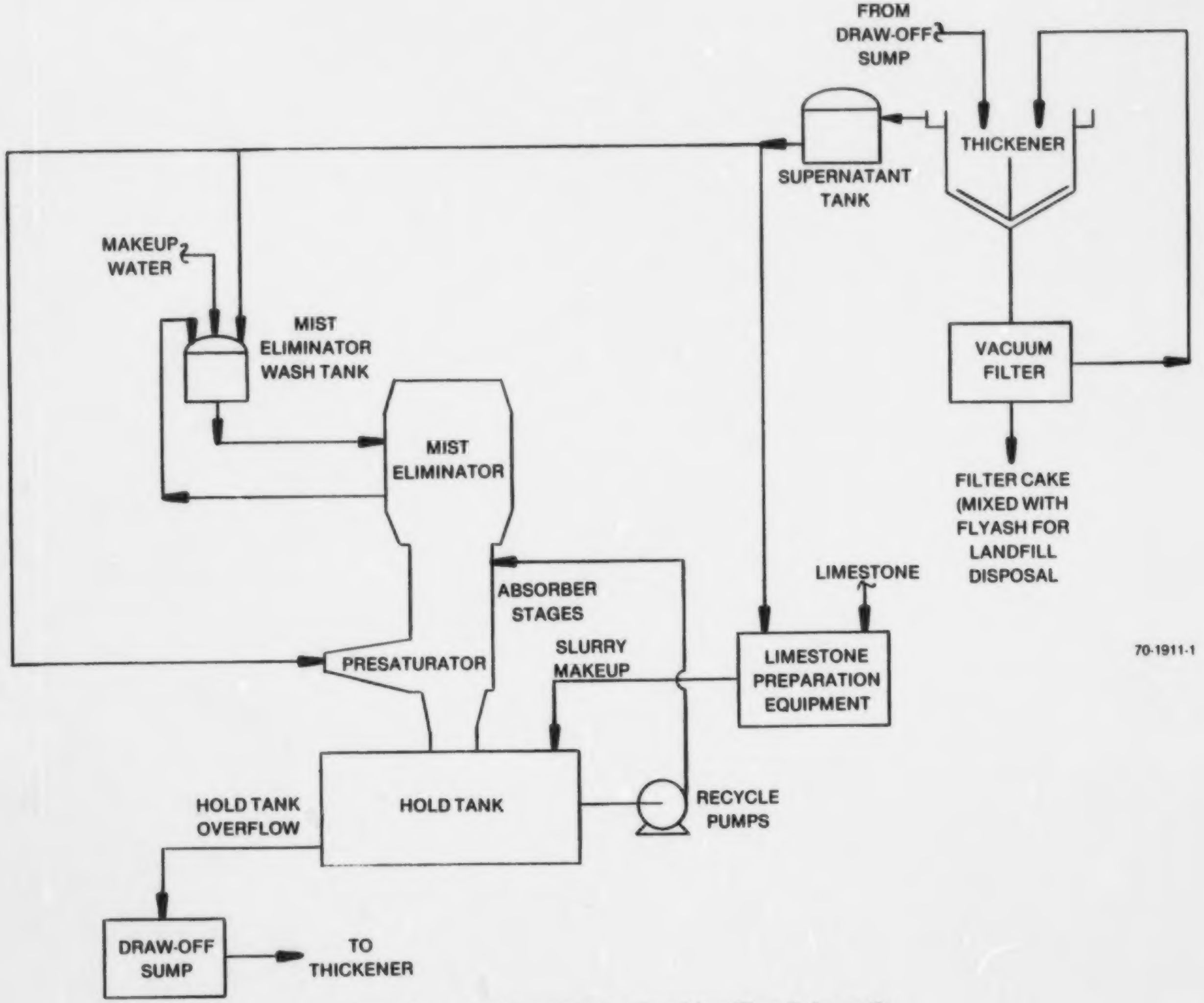

FROM DRAW-OFF SUMP
THICKENER
SUPERNATANT TANK
MAKEUP WATER
MIST ELIMINATOR WASH TANK
VACUUM FILTER
MIST ELIMINATOR
FILTER CAKE (MIXED WITH FLYASH FOR LANDFILL DISPOSAL
LIMESTONE
ABSORBER STAGES
SLURRY MAKEUP
LIMESTONE PREPARATION EQUIPMENT
PRESATURATOR
70-1911-1
HOLD TANK OVERFLOW
HOLD TANK
RECYCLE PUMPS
DRAW-OFF SUMP
TO THICKENER

Figure 2 Southwest Unit 1 Scrubber Flow Schematic.

This situation was improved by a redesign of the demister spray and presaturator spray systems as shown in Figure 3. Instead of a closed loop system, the demister spray system was changed to utilize supernatant water in a once-through flow sequence. This water was then collected and repumped to provide the source for the presaturator spray system, and to wash the underside of the trap-out-trays.

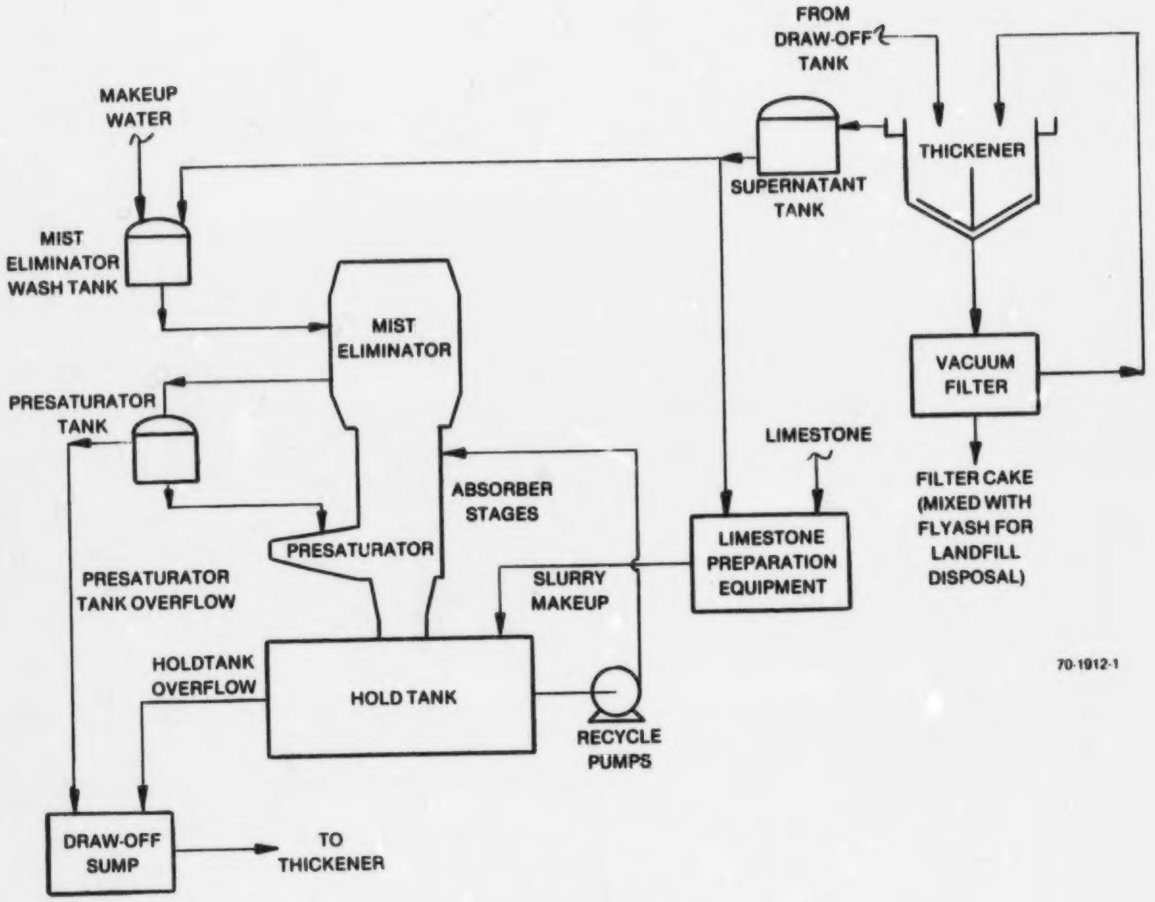
This modification has improved the operability with regard to the demisters. Pluggage and scaling occurs much less frequently, but improvements are still being sought by plant personnel.

Absorption area pluggage can be traced in part to the following factors: sphere failure, inadequate limestone grinding, and on-off-on cycling. Sphere failure has been a problem since the original system start-up. The original sphere supplied by UOP was a seven gram white (TPR) sphere similar in appearance to a ping-pong ball. The spheres were installed in only one layer per module initially. Two levels per module were added prior to the September, 1977 acceptance testing. It was evident after only a few weeks of continuous operation that the spheres were failing. Many ruptured and filled with slurry; others collapsed or dimpled losing their buoyancy. The sphere layers were no longer completely fluidized and behaved as solids filters resulting in severe absorber pluggage.

A random sampling of the spheres was conducted in September, 1978, to determine the failure rate of the TPR spheres. It was determined that virtually all of the spheres had either totally failed or were badly deformed. Following discussions with UOP it was decided to replace the TPR spheres with black foamed nitrile rubber spheres (eleven grams each). This replacement was made in October, 1978, with a bed thickness of approximately 8 inches as prescribed. Upon scrubber start-up after this sphere replacement, a significant increase in pressure drop through the scrubbers was detected. Only about ninety percent of full load could be reached because of the inability of the I.D. fans to make up the additional pressure drop. Additionally the spheres began absorbing moisture, thus reducing their buoyancy and creating the same type of pluggage problem that had existed with the damaged TPR spheres.

In an effort to solve this situation, the sphere bed depths were reduced from 8 inches per layer to approximately four inches per layer. Some increase in load carrying capability was realized for short periods of time, but the failure rate of the spheres was still rapid. Many split in two; others shriveled and cracked and lost their buoyancy. Weekly cleaning of the ball cages was sometimes not sufficient to prevent complete pluggage of the sphere layers. It was not uncommon for plant employees to dig spheres out of the pluggage with screwdrivers in attempts to clear the residue from the cages.

In early 1980, it was decided to evaluate other spheres for possible replacement. A sphere, manufactured by Puget Sound Trading Co., was selected for testing. These spheres are green in color and approximately the same size and weight as the original TPR spheres, but with cast ridges for additional structrual strength. The spheres were installed during the summer of


FROM
DRAW-OFF
TANK
THICKENER
SUPERNATANT
TANK
MAKEUP
WATER
MIST
ELIMINATOR
WASH TANK
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ELIMINATOR
VACUUM
FILTER
PRESATURATOR
TANK
LIMESTONE
FILTER CAKE
(MIXED WITH
FLYASH FOR
LANDFILL
DISPOSAL)
ABSORBER
STAGES
PRESATURATOR
LIMESTONE
PREPARATION
EQUIPMENT
PRESATURATOR
TANK OVERFLOW
SLURRY
MAKEUP
HOLDTANK
OVERFLOW
HOLD TANK
RECYCLE
PUMPS
70-1912-1
DRAW-OFF
SUMP
TO
THICKENER

Figure 3 Southwest Unit 1 Scrubber Flow Schematic.
Revised 5-77

1980 and to date no substantial pluggage problems have been encountered. Unit load-carrying ability has returned to original design levels.

City Utilities is currently evaluating the conversion of the scrubbers from fluidized bed contacting to a tray type design. This conversion has been completed at other locations and appears to have improved both the economics and the availability of the scrubbers.

Funds to provide and install limestone classification equipment following the ball mills have been budgeted. When this equipment is installed and put into service, it is anticipated that limestone utilization will improve and pluggage frequency decline.

Expansion Joints

The I.D. fan outlet expansion joints have been a source of considerable scrubber downtime and expense. The joints originally installed by UOP were manufactured of high-strength low-alloy steel. Within a few months of operation, it was evident that the joints were failing. During the fall 1977 outage, the steel expansion joints were replaced by UOP with Viton rubber joints. During 1978, over 3,000 hours of scrubber module downtime resulted from numerous failures of these expansion joints.

In early 1979, plant maintenance personnel accepted the responsibility of maintaining the joints from UOP. Joint life at that time could be expected to range from two hours to two or three weeks. An analysis of samples taken from failed joint specimens indicated an internal abrasion failure mode. The presence of hardened fly ash and limestone in the insulation boot on the flue gas side of the rubber was determined to be the source of the abrasion.

Further evaluation into the presence of the calcium material indicated that the probable cause was presaturator spray nozzle pluggage. With a nozzle plugged, flow of the spray could be directed into the duct counter to the flow of the flue gas stream.

In May, 1979, the expansion joint was redesigned. The Viton material was placed on the inside of the joint in contact with the flue gas. A neoprene belting material was utilized for an external cover with insulation fill between the two layers. Additionally, a small plate was installed across the floor of the ductwork downstream of the expansion joint to halt errant presaturator spray.

Since these modifications in mid-1979, the life expectancy of these expansion joints has increased to at least six months. It is anticipated that with slightly thicker belting material longer joint life can be achieved. The evaluation of expansion joint performance is ongoing.

Piping

Most of the scrubber piping systems at this installation are made of fiber reinforced plastic (FRP) pipe. The pipe, manufactured by Fibercast Co. (Div. of Youngstown Sheet & Tube Co.) and installed by UOP is resistant to abrasive wear.

The Southwest Unit 1 Scrubber installation is not enclosed; i.e., all piping systems are exposed to ambient weather conditions. The first winter of scrubber operation demonstrated the vulnerability of FRP pipe to cold weather failure. There were three types of failures: failure due to freezing; failure due to pipe embrittlement; and failure of the joint adhesive. Proper heat tracing and insulation of an FRP piping system is most difficult because of its poor conductivity. There were instances of FRP pipes with heat tracing and insulation that froze during the winter of 1978-79.

After evaluating various repair possibilities of the FRP piping systems and researching other piping options, it was decided during the summer of 1979 to replace the mist eliminator trap-out-tray piping system with a lined steel pipe system, manufactured by Peabody Dore. This replacement was accomplished in October, 1979. The new piping system was heat-traced and insulated in a proper manner. This piping system, historically the most susceptible to freezing, did not sustain a single failure during the 1979-80 winter period.

Because of the improvements noted in the new piping system, it is planned to replace all FRP scrubber piping with a lined-steel piping system in the fall of 1980. With proper heat-tracing and insulation, pipe freezing and breakage problems should be greatly reduced.

Corrosion

As with most scrubber installations, corrosion causes continuing and extensive maintenance. Corrosion has caused deterioration of dampers, seal strips, ducts, linings, and exposed metal surfaces in the outplant area.

Very shortly after initial scrubber start-up it became apparent that material selections were not what they should have been. The chloride concentration in the scrubber slurry has been measured as high as 2000 to 3000 ppm. Entrained mist that was not removed by the demisters collected in the outlet duct, exposing the dampers, seal strips, and lining materials to the high chloride liquid. In addition, continued contact of the liquid with the flue gas resulted in a lowering of its pH to between 1.0 and 1.5 due to further SO_2 absorption. The combination of this high chloride low pH environment resulted in severe corrosion and rapid material deterioration. The original outlet duct lining, Rigiflake 485, began peeling approximately two weeks after initial scrubber start-up in April, 1977. Attempts were made to spot-patch the failed liner areas. In October, 1977 the entire outlet duct surface was cleaned and relined again with Rigiflake. Within a month of operation, the liner had again failed.

Evaluations by the owner and UOP representatives led to the selection of a different liner material for installation during a planned outage in October, 1978. Two variations of the liner material, Placite 4005 and Placite 4030, were used. After being in service for approximately one month, the duct liners were inspected and found to be in very good condition. After the second month of service, deterioration of the liner was quite advanced. In April, 1979, the Placite application contractor returned to the site to patch the failed areas and any other areas that were deteriorated.

After a few months of service, liner failure was again apparent. The failure appearance was in most cases that of a blister. The Placite would separate from the metal duct allowing corrosion of the metal beneath the Placite. It was determined that improper metal surface preparation was the probable cause of the failures.

During the month of October, 1979, the ducts were sandblasted to white metal, and relined with Placite. During the process, all deteriorated metal areas were patched and turning vanes replaced. Within three months of operation, the Placite lining had deteriorated sufficiently in certain areas to allow holes to be eaten through the one-quarter inch A-36 steel duct material.

At this time other liner materials are being evaluated. Hastelloy G (Cabot) appears to be a prospect for use but is so expensive that the budget will not allow its use. Resista-Flake by Corrosioneering has some applications which seem to have served marginally well. At the time of this writing, no final decision had been made as to the material to be selected for duct liner repairs in the fall of 1980.

Some gains have been made in the selection of materials for dampers and seal strips. The original scrubber inlet and bypass damper seals were of 304L stainless steel; the frames and blades were of A36 steel. Within one month of operation, failure was evident. UOP then replaced the seals with 316L stainless, the same material as the outlet dampers.

By the fall fo 1977 it was evident that the 316L material was not suitable for the pH and chlorides present. UOP replaced the inlet and bypass damper seals and the outlet dampers in December, 1977. The inlet and bypass damper seal material used was Inconel 625 Huntington Alloy; the outlet damper material used was Udeholm 904L, including seal strips.

To date, the Udeholm 904L outlet dampers have provided good service. Some slight seal strip deterioration is evident and will be corrected. The Inconel 625 inlet and bypass damper seals have not performed as well. The seal strips have been completely corroded away and the carbon blades and framework are badly deteriorated. An evaluation of materials to replace these dampers is underway at the time of this writing.

The originally installed presaturator lining, Precrete Grout, began failing soon after its application and before the scrubbers were placed in service. Severe cracking appeared as if it were shrinkage induced. The installation contractor made repairs on two different occasions in an attempt to save the liner but to no avail. By late summer, 1977, holes were eaten through the outer wall of the duct.

During the outage that began in September of 1977, UOP removed the Precrete material and relined the entire presaturator area of the scrubbers with Udeholm 904L. This material served appreciably better than the Precrete. The Udeholm lining required no maintenance until October, 1979, when several sections had to be patched. Some degree of deterioration was evident over most of the Udeholm lining indicating that a more resistant material was needed. All repairs utilized Hastalloy G material in the presaturator area.

An inspection was conducted of the presaturator area during April, 1980. Further deterioration of the Udeholm lining was evident. Several materials were considered for repairs. Laboratory testing of Plastaloy (by Continental Alloy Steel) indicated it possessed high abrasion resistance and high corrosion resistance. It was decided to try some of this material in one scrubber module. The sheets of the material were installed using special nylon bolts. The material lasted less than one week in operation. Apparently the expansion of the material differed substantially from the metal causing the Plastaloy to twist and wrench its way apart from the duct. Investigation is continuing into materials for future use in this area.

Instrumentation

Many of the problems initially encountered in the instrumentation area were caused by long periods of inactivity while UOP performed needed modifications on the FGD Systems. When UOP left the job site City Utilities found itself without an adequately trained technical force to trouble-shoot and maintain the systems. The maintenance staff has been expanded and training provided so that we now have good capabilities to deal with instrumentation and control problems.

Freezing problems have beset many of the instrument systems since the first winter of operation. Instrument air drying capacity, as originally installed, was sorely inadequate. The passage of this inadequately dried air through small-diameter air control lines caused condensation in the air lines and eventual freezing. Damage to transmitters and various other instrumentation resulted. From initial scrubber operation until February, 1979, some 1500 hours of module downtime had occurred because of icing in air lines and instrumentation.

A new instrument air dryer installation was funded by City Utilities and the installation completed in February 1979. This dryer unit was of the dessicant type and has served well. During the spring and summer of 1979, the air lines and instruments were cleaned and purged to insure that no moisture remained in the lines. As a result, there were no instrument air line freezing problems during the winter of 1979-80. One plant air line which supplies air to the limestone ball mill clutch control did freeze last winter. A dessicant dryer assembly will soon be added to that air system to correct the situation.

Another instrument freezing problem which has caused considerable difficulty and module downtime is the pH monitoring system. Each scrubber module has an on-line pH analyzing system with two glass electrode sensors. The sensors are located in a small open tank through which the slurry flows continuously. During extremely cold weather, ice forms around the sensors usually resulting in breakage of the sensors. Pluggage of the sensors by a build-up of slurry requires frequent cleaning which eliminates the possible use of a full enclosure. New in-line pH cells are now on order.

Continuous monitoring of flue gas opacity and SO_2 has been totally unsuccessful. The original opacity monitors were Lear Sielger RM-4 instruments which were installed in the I.D. fan outlet ducts. They never operated reliably and failed soon after installation. The static pressure in the duct where they were installed could easily reach 20 inches (W.C.). The purge air fans on the instruments were inadequately sized to cope with this pressure, thus making cleaning and other maintenance of the monitoring equipment difficult because of flue gas infiltration. Corrosion of the equipment eventually rendered it useless.

The original flue gas SO_2 monitoring equipment was a model 1268-2-21 Turnkey Gas Analysis System installed by Dynasciences Inc. This was an extractive system with the sample obtained from the wet gas stream at an upper stack platform (255 feet above ground level). The sample line, heat traced and insulated, ran from this sample point down to the ground, then to the scrubber control room where the monitoring equipment was located. The total length of this sample line was over 400 feet. During operation, sample line pluggage was nearly a daily occurence. If the sample probe plugged, a technician was required to climb the stack ladder to the sample probe level to clean and repair the probe. This was quite an unpopular duty, and during wet or freezing weather was unsafe.

After discussions with various vendors, and continued lack of success with further equipment modifications, it was determined that the existing systems were not capable of operating in a reliable manner.

City Utilities evaluated consultants that specialized in the field of flue gas monitoring and testing and selected Entropy Environmentalists, Inc. to study the problems of the existing installations and to evaluate modifications or redesign of the monitoring equipment installation. Some of this work has now been completed. A new, low pressure zone location has been tentatively selected for the opacity monitors. The monitors would in fact be located at the I.D. fan inlets in a negative pressure zone. The flue gas SO_2 sampling point will be relocated to a lower stack platform just above the scrubber outlet duct. A stairway is planned to connect the scrubber to this platform and eliminate the need to climb the stack ladder to service the equipment.

The specific equipment to be used is still being evaluated. Performance specifications have been prepared. Entropy is not presently aware of any monitoring equipment vendor who has successfully installed a system on a wet stack which operated as reliably as required by the involved governmental agencies. Few, if any vendors are willing to warranty their equipment installation for any period of time after an acceptance test and their people leave the site. Our experience has been that a vendor's serviceman can get his equipment in operation, leave the site, and his equipment would again be inoperative before he would reach the airport.

Thickener/Dewatering Systems

The waste slurry enters the thickener from the scrubbers with between five and fifteen percent by weight solids content. Solids settling in the thickener is aided by the addition of polymer. Frequent sampling of the mixture to determine the settling rate is required to maintain proper interface levels. The solids in the slurry must not settle out to more than thirty to forty percent by weight or pluggage of lines and pumps occurs. If the solids stay in suspension they overflow the thickener and are returned as supernatant to the scrubber. These returned solids can contribute to pluggage of demister nozzles.

One problem encountered in the thickener tank has been anaerobic bacterial attack on sulfites and sulfates resulting in a change of color of slurry and cake from cream to gray with a resultant smell of rotten eggs (H_2S formation). This problem occurs during warm weather and when the slurry is retained in the thickener during breakdowns on the system lasting longer than a day. The bacteria have been controlled by "shocking" the thickener tank contents with swimming pool grade granular chlorine (usually four-hundred to five-hundred pounds broadcast into the thickener tank that has 750,000 gallons capacity).

The under flow from the thickener is pumped to an EIMCO Vacuum Filter belt and discharged to conveyors with sixty to seventy percent solids. A conveyor transports the dewatered sludge to a pug mill where it is mixed 50/50 (by weight) with dry fly ash. This produces a material which is directly landfilled. During freezing weather, spillage of the material causes pluggage and freezing of the conveyor tracks. Torn conveyor belts, which results in temporary shutdown of the system, have been common. Complete enclosure of this process has been comtemplated but funds have not been available to accomplish the work.

A limiting factor in the dewatering operation is the pug mill. Its capacity is such that at continuous high operating levels, sixteen to twenty hours of operation per day are required to maintain the solids level within the thickener tank at an acceptable level.

There is no redundancy of conveyors or the pug mill. If one breaks, the system is down and the draw-off from the thickener must either be pumped to the emergency pond or discharged to the ash pond. This type of breakdown would not immediately affect the operation of the scrubbers.

Other

Beneath each scrubber module in the hold tank is a large turbine agitator manufactured by Lightnin Co. The drive shaft for the agitator is approximately twenty-two feet (22') in length and is six inches (6") in diameter. The shaft in the "B" scrubber module has now fractured twice. The first break occurred in May, 1979. An analysis of the break indicated frequency related flexural fatigue failure. While awaiting replacement material, the design of the hold tank and agitator shaft was evaluated to determine the cause of flexure in the shaft. No conclusive determination was made so a new shaft was installed and the scrubber returned to service in July, 1979. This new shaft failed in May, 1980. This failure differed from the initial one in that it occurred at a factory weld and there was no indication of fatigue. A new shaft was fabricated locally and the scrubber returned to service. The cause of this failure is still being diagnosed by the manufacturer.

Consideration has been given to two other additions which would undoubtedly improve overall scrubber availability. They are:

a) a spare (3rd) scrubber module
b) freeze-proof the FGD systems by totally enclosing them.

The cost of providing these improvements have prohibited serious consideration of them.

ADIPIC ACID TEST PROGRAM

City Utilities is a progressive organization and is interested in improving its operation. When it was learned that the Environmental Protection Agency (EPA) was proposing to sponsor a full scale demonstration of adipic acid addition to limestone scrubber operations, management was interested.

A contract now exists between Radian Corporation and EPA for this demonstration program. City Utilities is providing the host site (Southwest Power Station Unit No. 1) and other support services. The Air Correction Division (ACD) of UOP is also participating in the program.

The addition of weak organic acids such as adipic acid to limestone FGD systems has been shown to benefit both SO_2 removal and limestone utilization and, also, to have a potential for improving the overall operability of a limestone FGD system. Adipic acid has the effect of buffering scrubber solutions, thereby enhancing liquid phase mass transfer. EPA has tested adipic acid addition to limestone scrubbers at a 0.1 MW pilot plant in Research Triangle Park, North Carolina and at the 10 MW prototype units at Tennessee Valley Authority's Shawnee power plant near Paducah, Kentucky with encouraging results. The program at Southwest Unit 1 is the final step in demonstrating adipic acid as an additive for commercial FGD systems. At the time of this writing, the demonstration program was in the second month of a scheduled six-month program.

Test Program Objectives

EPA objectives in this program are to confirm the results of their previous testing and successfully transfer this technology from the pilot and prototype stages to a full-scale limestone FGD system operating in both a forced oxidation and natural oxidation mode. City Utilities Southwest Unit 1 represents a nearly ideal system for accomplishing these objectives for the following reasons:

the station fires high sulfur bituminous coal;

the FGD system already includes a thickener, a vacuum filter, and a clay-lined landfill which are typical of current dewatering and waste disposal techniques;

the potential exists to increase SO_2 removal, improve limestone utilization, and increase the efficiency of the dewatering train;

the unit is a commercial scale power plant that will require relatively minor modifications to perform the test program, thereby resulting in the most efficient use of EPA funds.

City Utilities expects to gain valuable information on the operation and performance of its scrubber. Its objectives in the program are three fold:

aid in the successful completion of the demonstration program by providing the host site;

evaluate the operating and cost advantages and disadvantages of adipic acid addition and forced oxidation at the SWPP scrubber;

investigate the ability of adipic acid addition to keep the unit within compliance with the SO_2 New Source Performance Standards.

Anticipated advantages that City Utilities will see in their scrubbers operation following adipic acid addition include:

increased SO_2 removal, and

improved limestone utilization.

Increased SO_2 removal has several potential benefits with respect to operation of the scrubbers. First, the pH of the liquid in the outlet duct should show a substantial rise above the normal range of 1.0 to 1.5. In fact, a sample of this liquid has been tested during the recent preliminary adipic acid testing and its pH has been found to have increased to 3.7. This is due to the lower gas phase SO_2 concentrations in the outlet duct. Liner and duct corrosion rates should therefore be decreased for this reason. In addition, potential exists for removing some of the ball charge in each scrubber and still having high enough removal to keep the unit in compliance. Removing balls would have two positive affects. First, the pressure drop through the scrubbers would be decreased resulting in lower power costs for the I.D. fans. Secondly, a smaller ball charge would mean less chance for pluggage due to ruptured or deformed balls; thereby improving scrubber availability.

The increased utilization of limestone is also expected to result in certain system operating improvements. First, there should be less dissolution of limestone on the TCA beds and demisters, therefore decreasing the potential for scaling. Also, fluctuations in SO_2 removal efficiency caused by variations in routine process conditions may be reduced due to the buffering capacity of the adipic acid. These advantages will be compared to the additional operating cost incurred by adding the adipic acid.

Participants In Demonstration Program

The Environmental Protection Agency has funded the development work on adipic acid as an additive to limestone FGD systems. This program represents the agency's final step in upgrading the technology to commercial status. John Williams is acting as the project officer for the EPA Industrial Environmental Research Laboratory (IERL) at Research Triangle Park (RTP) in this demonstration program. The EPA contracted Radian Corporation to conduct the test program and provide the necessary FGD expertise to evaluate the program results. Radian has subcontracted City Utilities of Springfield, Missouri and UOP's Air Correction Division to provide the test site and support for the necessary system modifications.

As prime contractor, Radian's responsibilities in this demonstration effort include overall project management and coordination, conceptual design of the forced oxidation and adipic acid feed systems, development and implementation of the test program, evaluation of results, and reporting.

Air Correction Division will prepare a detailed design and specifications for the forced oxidation and adipic acid feed systems; review quotes and select vendors; and procure, install, and start-up the involved equipment.

The primary responsibilities of City Utilities in this demonstration program include power plant and scrubber operation, review of proposed site modifications, coordination of site modifications and interfacing with both Radian and UOP during onsite testing.

Proposed Test Plan

The adipic acid demonstration program will be divided into two test phases. The first phase will be a series of tests in the natural oxidation mode (present equipment configuration). Prior to the second test phase, the system will be modified so that air can be introduced into the reaction tank for forced oxidation testing in Phase II.

Within each of these phases, a one-month duration test without adipic acid (baseline test) will be conducted followed by two months of testing with adipic acid. Thus, the total program duration will be six months: three months in the natural oxidation mode and three months in the forced oxidation mode of operation.

The current plan is to conduct this baseline natural oxidation test during August, 1980. This will document the manner in which the system presently operates. The scrubbing system will follow boiler load during this period and several fairly short-term tests will be conducted at various pH levels. However, most of this month will be used to monitor the system's operation as City Utilities normally operates it.

Following this initial baseline test, adipic acid will be added to the system, and the effects on system performance will be monitored. Two weeks of adipic acid testing are planned between September 1st and 15th, when a one month scheduled outage will begin. An additional six-weeks of testing with adipic acid in the natural oxidation mode of operation will be conducted after the outage. The forced oxidation baseline testing should be conducted in December, 1980, with two months of adipic acid-forced oxidation tests planned for January and February, 1981.

The adipic acid will be added to the limestone sump with the limestone from the ball mills as shown in Figure 4. An on-off controller tied to the limestone feed rate to the ball mills will insure that the desired amount of adipic acid is maintained in the reaction tank. This desired adipic acid concentration can be altered by changing the flow rate from a weigh feeder. Adipic acid inventory in the weigh feeder will be maintained from a system consisting of a vibratory hopper and screw feeder. The adipic acid concentration in the reaction tank slurry will be analyzed periodically to insure that the desired adipic acid concentration is being maintained.

The oxidation air will be introduced through a sparger network consisting of PVC pipe in the reaction tank. The sparger pipe will be located fairly close to the walls of the rectangular reaction vessel to minimize chances of damaging the agitator shaft. The sparger ring will be installed during the September outage to minimize downtime. Four 1800 SCFM compressors will be utilized to supply air to both reaction tanks. An oxygen to SO_2 sorbed stoichiometry greater than 2.5 can be maintained at full load with this air rate.

Since this program is a demonstration program rather than a research program, only a minimum of parametric testing will be performed. However, changes in scrubber feed pH, adipic acid concentration, and air/SO_2 stoichiometry will be made to find the optimum operating conditions within each test phase. Optimum performance will be evaluated by examining such parameters as SO_2 removal efficiency, limestone utilization, required adipic aicd feed rate (unaccounted for losses of adipic acid), and sludge dewatering properties.

Preliminary Test Results

The initial results of the adipic acid testing in early September were very encouraging. Prior to the addition of adipic acid to S-1 module, its SO_2 removal efficiency had averaged about 65 percent at the normal operating pH of 5.5. After addition of adipic acid to a liquid phase concentration of between 800 and 1000 ppm, the removal efficiency of the module increased to above 90 percent with a high of 95 percent at high

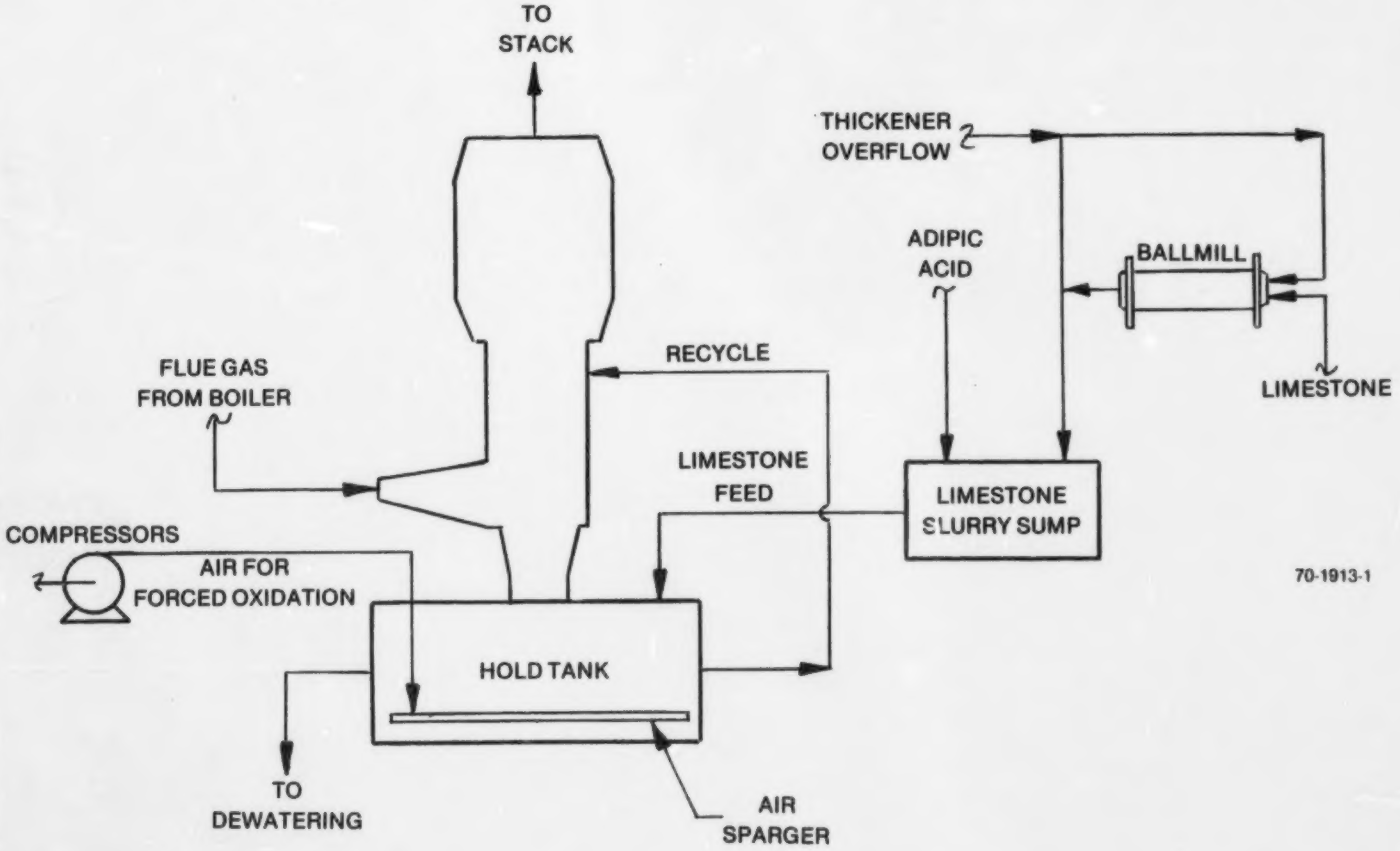
TO STACK
THICKENER OVERFLOW
ADIPIC ACID
BALLMILL
LIMESTONE
FLUE GAS FROM BOILER
RECYCLE
LIMESTONE FEED
LIMESTONE SLURRY SUMP
COMPRESSORS
AIR FOR FORCED OXIDATION
HOLD TANK
TO DEWATERING
AIR SPARGER
70-1913-1

Figure 4 Southwest Unit 1 Adipic Acid Test Program Flow Diagram.

load conditions. Improvements in limestone utilization were also noticed. Results of a later test at an operating pH of 5.0 and an adipic acid concentration of 1500 ppm showed greater than 90 percent SO_2 removal with limestone utilization of 99 percent. Testing is scheduled to begin again after the maintenance outage.

CONCLUSIONS

City Utilities is continuing its efforts to improve the availability and reliability of Southwest Unit 1 FGD Systems. Corrosion and absorber area pluggage remain to be major problems to overcome.

The use of adipic acid as a limestone additive appears to provide an interesting alternative to be considered while evaluating improvements in system operation. Improved limestone utilization and SO_2 removal will provide an economic basis for comparison with other potential alternatives.

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RESULTS OF THE CHIYODA THOROUGHBRED-121 PROTOTYPE EVALUATION

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ABSTRACT

A ten-month evaluation of the Chiyoda Thoroughbred 121 Prototype ee Process (CT-121) was conducted at the Sholz Electric Generating Station of Gulf Power Company. The 23-megawatt CT-121 prototype was modified from existing CT-101 process equipment at Scholz by Chiyoda International Corporation, a subsidiary of Chiyoda Chemical Engineering and Construction Company, Ltd. Chiyoda operated the prototype, and the Electric Power Research Institute and Southern Company sponsored technical evaluations of the prototype process performance. This paper summarizes the findings of these evaluations. Detailed results of the gypsum stacking evaluation will be presented with the Chiyoda Thoroughbred 121 presentation.

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RESULTS OF THE CHIYODA THOROUGHBRED-121 PROTOTYPE EVALUATION

INTRODUCTION

The Flue Gas Desulfurization Program area of the Electric Power Research Institute (EPRI) is charged with responsibility of identifying, evaluating, and advancing FGD technology to help the electric utility industry meet current sulfur dioxide emission standards in the most efficient, reliable, and economic manner. The Chiyoda Thoroughbred-121 (CT-121) system was reported to offer technical and economical advantages over currently available flue gas (FGD) desulfurization technology. As a result, the EPRI and Southern Company Services sponsored a program to have Radian Corporation of Austin, Texas evaluate the Chiyoda Thoroughbred-121 (CT-121) process. (The Southern Company is an electric utility holding company operating in the Southeast. It includes Alabama Power Company, Georgia Power Company, Gulf Power Company, Mississippi Power Company, and Southern Company Services, Inc.) As part of this program, Chiyoda International Corporation, the American subsidiary of Chiyoda Chemical Engineering and Construction Company of Japan, built and operated a prototype (23 MW) CT-121 process at Gulf Power Company's Scholz Power Station near Sneads, Florida with the cooperation of Gulf Power Company. To a large extent, this system was constructed by modifying the existing CT-101 demonstration equipment at Scholz.[1] The CT-121 process at Scholz is designated as a prototype because it was the first large-scale application of the CT-121 process.

SYSTEM DESCRIPTION

Figure 1 shows a schematic of the Scholz prototype CT-121 plant. This system was designed to treat 53,000 standard cubic feet per minute (85,000 normal cubic meters per hour) of flue gas (23 megawatts of electrical production at Scholz). However, during the test program, gas flows ranging from 25,000-55,000 scfm were studied.

As shown in the figure, the inlet gas was cooled and saturated liquid stream in a venturi before entering the met bubbling reactor (JBR) where the bulk of the SO_2 removal occurred. Compressed air was injected into the JBR to completely oxidize

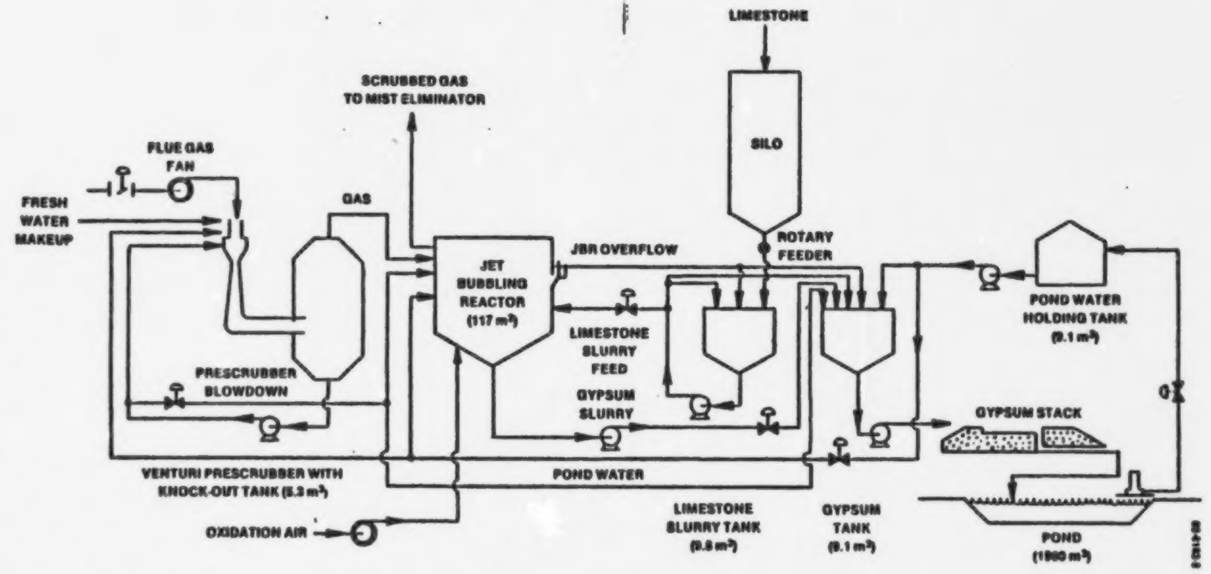

LIMESTONE
SILO
ROTARY FEEDER
SCRUBBED GAS TO MIST ELIMINATOR
FLUE GAS FAN
FRESH WATER MAKEUP
GAS
JET BUBBLING REACTOR (117 m³)
JBR OVERFLOW
LIMESTONE SLURRY FEED
GYPSUM SLURRY
PRESCRUBBER BLOWDOWN
VENTURI PRESCRUBBER WITH KNOCK-OUT TANK (5.3 m³)
POND WATER
OXIDATION AIR
LIMESTONE SLURRY TANK (9.8 m³)
GYPSUM TANK (9.1 m³)
POND WATER HOLDING TANK (9.1 m³)
GYPSUM STACK
POND (1900 m³)

Figure 1. Simplified process flow diagram of Scholz prototype CT-121 flue gas scrubbing system.

the sorbed SO_2 and to assist the agitator in maintaining a good gypsum solids suspension in the slurry. From the JBR, the gas passed through a mist eliminator prior to exiting the system through a glass reinforced plastic stack. There was no reheat provision in the prototype system. Powdered limestone was slurried and added to the JBR to control pH. Limestone grinding facilities were not included in the prototype. The gypsum produced during the evalution program was diposed of in a gypsum stack, a disposal technique commonly used in the phosphate fertilizer industry. Basically, a gypsum stack is a free standing body in which solid-liquid separation is achieved by solids settling in a hollowed out section on top of the stack. The supernatant liquid flows through the walls of the stack to form a "moat" around the stack. This disposal was evaluated independently by Ardaman & Associates of Orlando, Florida under EPRI research project 536-3 during the CT-121 demonstration. (A copy of the report was distributed at the EPA symposium as an unpresented paper.)

The unique and central feature of the CT-121 process is the jet bubbling reactor. Figure 2 shows a schematic of the prototype JBR configuration. SO_2 removal, sulfite oxidation, limestone dissolution, and gypsum crystallization reactions are all accomplished within this one vessel. This concept deviates from the conventional limestone system which contains large recycle pumps, separate absorption vessels and reaction tanks. Such a scheme can affect the capital cost of a FGD system. In the JBR, the gas is dispersed several inches beneath the slurry. This minimizes the liquid phase mass transfer resistance which can limit SO_2 removal in conventional spray tower systems. The liquid pumping power requirements are also low in the CT-121 system because large slurry recirculation pumps are not used, however, the power required to overcome the high gas side pressure drop tends to offset this savings somewhat. Figure 3 shows the physical arrangement of the JBR, the inlet and outlet ducts and mist eliminator at Scholz.

TEST PLAN AND OBJECTIVES

The objective of the program was to evaluate the performance of the CT-121 system under a wide range of operating conditions and to measure the reliability of this prototype. By varying both site-specific and some non-site-specific parameters, an "operating envelope" in which the CT-121 system can function successfully was determined. This performance evaluation therefore provides a basis for cost evaluation activities as well as for some of the design parameters required for commercial units.

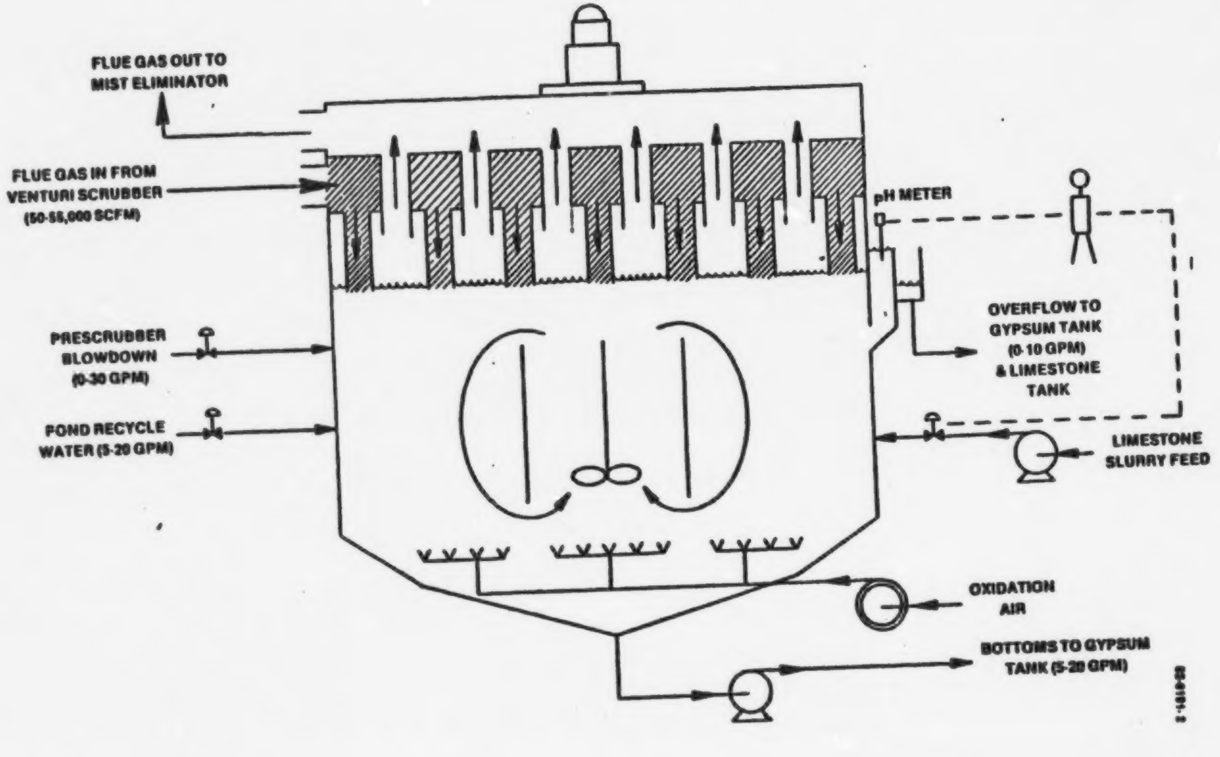

FLUE GAS OUT TO MIST ELIMINATOR
FLUE GAS IN FROM VENTURI SCRUBBER (50-55,000 SCFM)
PRESCRUBBER BLOWDOWN (0-30 GPM)
POND RECYCLE WATER (5-20 GPM)
pH METER
OVERFLOW TO GYPSUM TANK (0-10 GPM) & LIMESTONE TANK
LIMESTONE SLURRY FEED
OXIDATION AIR
BOTTOMS TO GYPSUM TANK (5-20 GPM)
80-8191-2

Figure 2. Schematic of jet bubbling reactor (JBR).

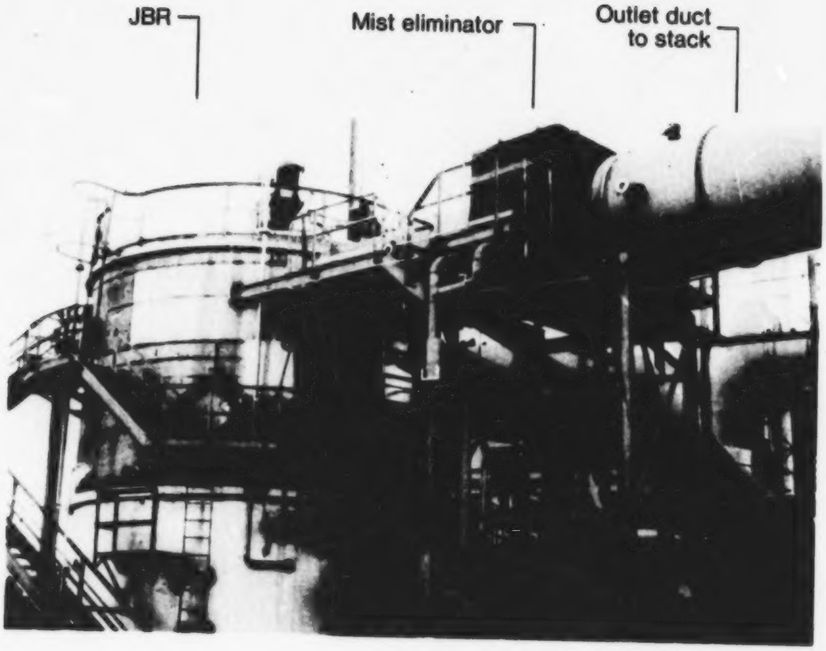

JBR
Mist eliminator
Outlet duct to stack

Figure 3. Jet bubbling reactor, gas ducts, and mist eliminator.

The testing of the prototype CT-121 system was divided into four phases:

- Phase 0 - Three-month duration startup, shakedown, and initial parametric tests conducted by Chiyoda.
- Phase I - Two-month duration baseline tests conducted at Chiyoda specified operating conditions to quantify some of the control variable fluctuations that might be encountered during routine operation.
- Phase II - Four-month duration test series conducted under a variety of operating conditions (forced variable perturbations) to evaluate system response under operating conditions that may be representative of a broad scope of utility applications.
- Phase III - Three-week duration tests conducted by Chiyoda following modifications to the JBR internals to simplify the JBR design and reduce capital cost.

In all, a total of ten months of tests were conducted over an eleven-month period beginning in August 1978. Throughout the program locally hired personnel, operated the system. Chiyoda provided supervision only during the day shift. Chiyoda conducted Phase 0 with no input from EPRI although operating data were transmitted to EPRI. Phases I and II constituted the EPRI evaluation program. During these phases, the test conditions were proposed by Radian and approved by EPRI, SCS, and Chiyoda; an on-site Radian test crew conducted the tests and reviewed operating conditions with Chiyoda personnel. During Phase III, Chiyoda performed the testing independently, but Radian observed the testing as EPRI's and SCS's representative.

TEST RESULTS

Synopsis

When judged by five critical performance criteria: SO_2 removal efficiency, solid waste quality, limestone utilization, resistance to chemical scaling and reliability, the performance of the CT-121 process throughout the EPRI evaluation program was quite good. SO_2 removal efficiencies of 95 percent with an inlet flue gas concentration of 3500 ppm SO_2 were achieved, and the gypsum produced throughout the program settled rapidly and dewatered easily. The operation of the prototype system was particularly outstanding from the standpoint of limestone utilization and chemical scale control. Limestone utilization within the JBR averaged over 98% for the evaluation program. A detailed inspection at the conclusion of Phase II revealed only minimal chemical scale deposition, none of which posed a significance operating problem. This was after nine months of testing including three months of Chiyoda shakedown operation and six months of EPRI-sponsored tests. These performance results are excellent in view of test conditions which deviated significantly from Chiyoda's design operating set points. These results thus indicate that the system is flexible and can withstand significant process upsets. These results demonstrated that Chiyoda prototype CT-121 is an FGD capable of continuous, reliable, and efficient operation.

SO_2 Removal. The JBR overflow pH and JBR WP influenced the SO_2 removal efficiency to the greatest extent in the test program. The SO_2 content of inlet flue gas showed a marked effect on removal efficiency only at concentrations above 2200 ppm. The oxidation air stoichiometry and flue gas flow rate altered SO_2 removal characteristics of the JBR chloride levels up to 6000 mg/l did not have a measurable effect on SO_2 removal efficiency.

Three parameters, pH, WP, and inlet SO_2 concentration were fit to a theoretically derived expression for SO_2 removal efficiency. The basic form of the mathematical equation was initially derived by Chiyoda in 1978. Tests varying sulfur dioxide (O/SO_2) stoichiometry and gas flow were fairly short term in nature and were not varied in conjunction with variations in other process conditions. Because of this these were not included in the mathematical model for predicting SO_2 removal.

The 229 data points used for this analysis were best fit by using two equations, the first for inlet gas SO_2 concentrations less than 2200 ppm and the second for higher SO_2 levels. The primary reason for using two equations is that nearly all the testing was at an SO_2 concentration less than 2200 ppm (200 data points). A single equation predicted accurate results for SO_2 levels less than 2200 ppm but did not adequately predict the removals observed at higher SO_2 concentrations. Therefore, another set of equation coefficients were determined to better fit the data at higher SO_2 concentrations.

Equation 1 predicts the removal for inlet SO_2 levels less than 2200 ppm while Equation 2 describes the results achieved at the higher concentrations for the prototype CT-121 system.

Fractional SO2 removal =
(for inlet gas SO2
concentrations less
than 2200 ppm)

$$\frac{1 - \exp\left(-3.49 N \frac{WP}{15.4} 0^{1.07}\right|}{1 + 56.9\ N \frac{15.4}{WP} 0\ N 10^{-pH} 0 N \frac{SO_2}{1000} 0^{.11}} \qquad (1)$$

Fractional SO2 removal =
(for inlet gas SO2
concentrations greater
than or equal to 2200 ppm)

$$\frac{1 - \exp\left(-3.85 N \frac{WP}{15.4} 0^{1.44}\right|}{1 + .84\ N \frac{15.4}{WP} 0 N 10^{-pH} 0 N \frac{SO_2}{1000} 0^{5.26}} \qquad (2)$$

where WP is the JBR pressure drop expressed as inches of water, SO_2 is the inlet flue gas sulfur dioxide concentration in ppm and the pH is that measured at the JBR overflow. Both of the equations are applicable only to the range of Phase II test conditions at full load gas flow and with O/SO_2 stoichiometric ratios greater than 8. Figure 4 is a plot comparing the measured removal with the values generated by these two equations.

The equations show the importance of the pressure drop and SO_2 concentration pH, on removal. As the pressure drop increases, the exponential term decreases, thus predicting a higher SO_2 removal. Likewise, as the pH increases, increased SO_2 removal is predicted since both equations' denominators approach unity. Increases in either pH or WP were expected to improve SO_2 removal efficiency. Since increased pH results in decreased SO_2 back pressure in the froth zone, and increased WP reflects longer gas-liquid contact time and/or more efficient flue gas sparging. Figure 5 shows the effects of tradeoffs between WP and pH on SO_2 removal with the prototype unit. In most situations, it will be more desirable to obtain the required SO_2 removal by using higher pH due to the relatively low cost of limestone. This should be evaluated on a case by case basis, and caution must be used to ensure that increased limestone concentrations do not cause scaling problems.

Further examination of Equations 1 and 2 show that SO_2 concentration had minimal effect on SO_2 removal when the inlet SO_2 concentration remained below about 2200 ppm. Above this level, increases in SO_2 concentration caused a fairly rapid decline in SO_2-term exponent in Equation 2 and is shown in Figure 6. This

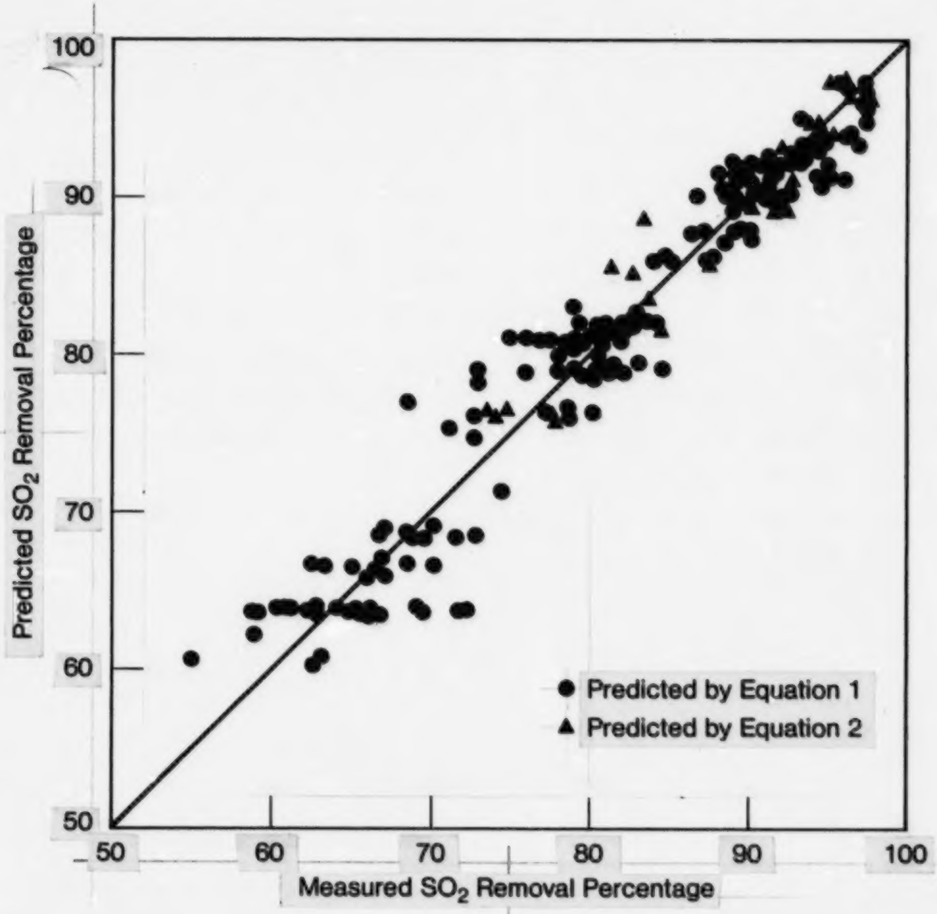


Figure 4. Predicted versus measured SO_2 values for Phase II data using Equations 1 and 2.

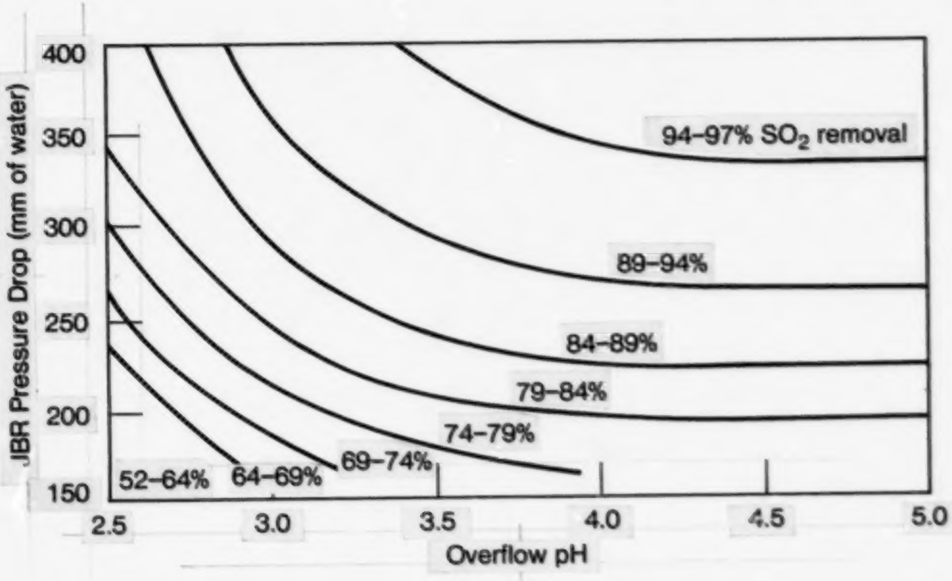


Figure 5. SO_2 removal percentage as a function of pH and pressure drop predicted by Equation 1, 2000 ppm SO_2 inlet concentration.

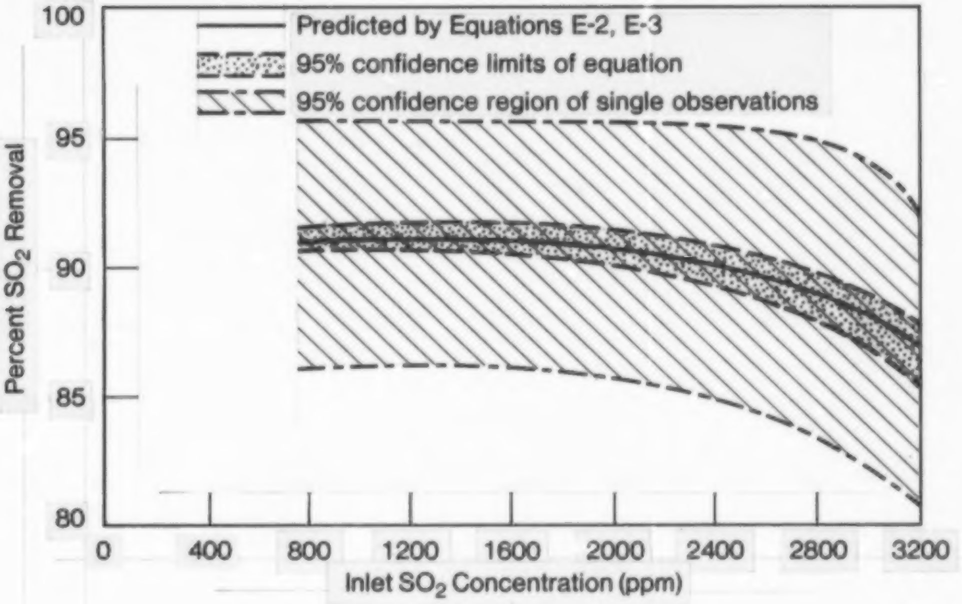


Figure 6. Fractional SO_2 removal versus input SO_2 with fixed pressure drop ($\Delta P = 11.5''$, pH = 3.5).

drop-off in SO_2 removal occurs at an SO_2 concentration higher than in many spray tower designs mainly because of the increased liquid surface renewal rate and increased interfacial mass transfer area created by the JBR design.

Even though the flue gas flow rate and the oxidation air rate were not included in the predictive equations, these variables had a measurable impact on the SO_2 removal rate. The boiler variable load tests in Phase I indicated that flue gas flows lower than 30,000 scfm resulted in an average removal of 94% from a flue gas containing 1000-1200 ppm SO_2 concentration. Flows of above 45,000 scfm during the variable load test period resulted in an average removal of 90 percent. Unfortunately, there was not sufficient time to test the impact of low gas flows at different pH's, WP's, and SO_2 concentrations.

The results of several short duration tests emphasized the importance of maintaining rapid oxidation to achieve good SO_2 removal in the JBR. These short-term tests quantified the effect of air-rate (stoichiometry) on SO_2 removal as shown in Figure 7. While no difference between air rates of 1000 and 1300 scfm (1600 and 2090 Nm^3/hr) (O/SO_2 stoichiometric ratios ranging from 8 to 11) was seen in the initial tests, Figure 7 shows a reduction in SO_2 removal efficiency to about 77 percent at an air rate of 480 scfm (770 Nm^3/hr) (O/SO_2 stoichiometry of about 4). With the air shut off, the SO_2 removal dropped to below 40 percent. In addition to the O/SO_2 stoichiometry, distribution of air in the JBR (which is influenced by such factors as air sparger, agitator performance and specific draft tube design) is also important in maintaining good sulfite oxidation efficiencies. These design factors were not examined in detail in this evaluation program.

Changes in limestone sources and increased chloride concentrations in the JBR slurry had no measurable effect on SO_2 removal. The main difference between the Southern Materials Company (SMC) limestone and the Georgia Marble limestone was the particle size since both were high calcium limestones. The specified SMC limestone grind was 90 percent through 200 mesh (74)m) and the specified Georgia Marble grind was 90 percent through 325 mesh (44)m). The driving force for dissolution was sufficiently high at the low operating pH's in the CT-121 prototype for limestone size to have no effect on SO_2 removal. Likewise, spiking the system with 6000 ppm chloride (added as $CaCl_2$) had no effect on SO_2 removal.

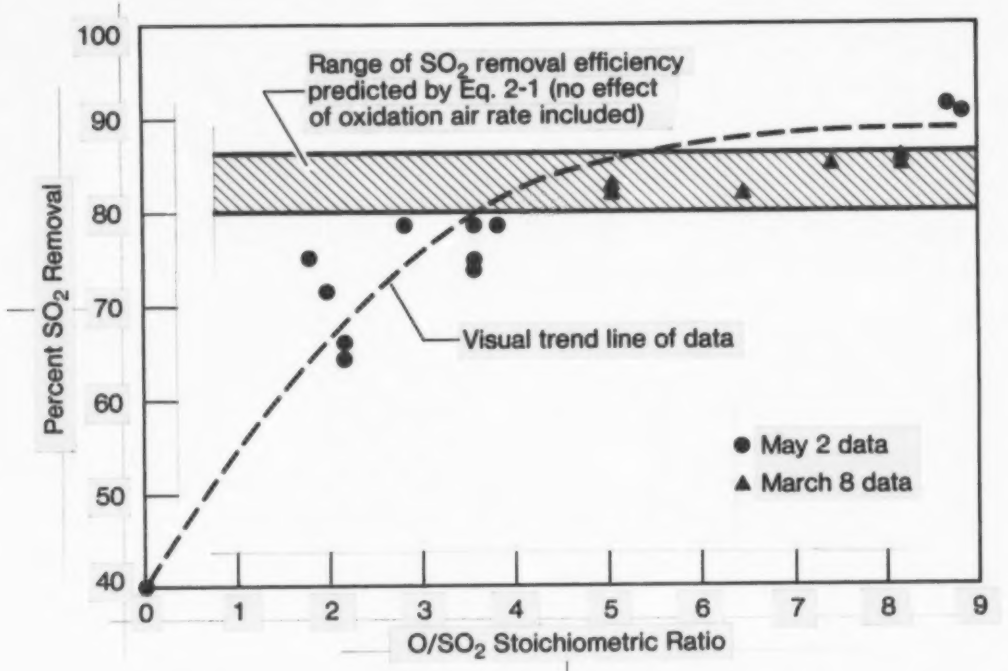


Figure 7. Low oxidation air test results — Phase II ($\Delta P = 9.0$ to $9.5''$ H_2O; pH = 3.2 to 3.8; inlet $SO_2 = 1400\text{–}1600$ ppm).

Changes in the JBR configuration made by Chiyoda prior to Phase III appeared to have only minor effects on SO_2 removal efficiency. Removal efficiencies of only one to two percent lower than those calculated in Phase I and II were calculated even though flue gas velocities through the sparger openings had increased 40 percent.

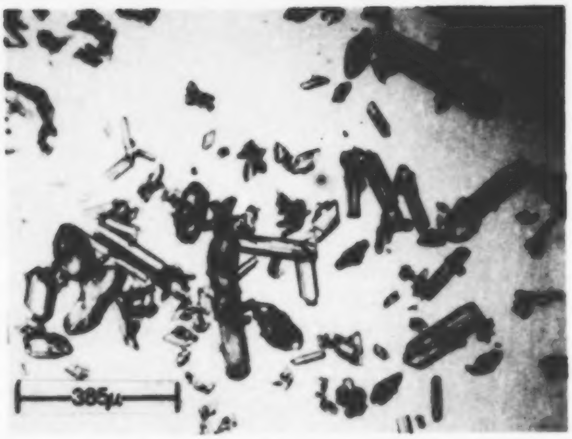
Limestone Utilization. Throughout the program the observed limestone utilization in the CT-121 system was quite high. For both Phases I and II, the utilization measured around the JBR remained above 98 percent. Changing the JBR overflow pH from 2.5 to above 4.5, and the limestone grind from 90 percent less than 325 mesh (44)m) to a grind of 90 percent less than 200 mesh (74)m) did not cause a measurable change in utilization.

The utilization was also good even when one considers the limestone added to the gypsum tank for final neutralization of the gypsum slurry to a pH of 6. Optimization of this process step was not an objective of the program. The limestone flow to the gypsum tank was only occasionally adjusted because there were no on-line pH monitors or controllers on the gypsum tank. The samples taken during Phase II indicated that the overall utilization including the neutralization tank, was somewhat lower in Phase II (f93 percent) than in Phase I (f97 percent). However, it appeared that the multiple changes in process conditions which occurred in Phase II may have caused some pH upsets in the gypsum tank. This was probably the primary cause of the lower utilizations.

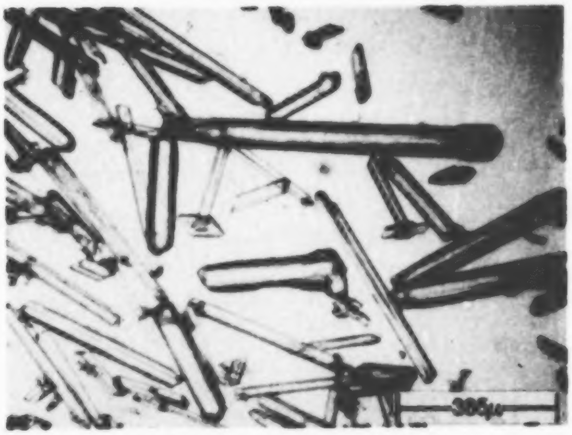
During Phase III, Chiyoda tested JBR overflow pH set points approaching six. At these conditions, the utilization in the JBr dropped to about 87 percent.

Solids Characteristics/Gypsum Scaling Tendency. The solids produced in the JBR during the evaluation were generally greater than 97% gypsum. There were no sulfite solids measured since the pH was always low enough that calcium sulfite would remain in solution until it was oxidized. Also, as discussed in the preceding limestone utilization section, there were only small amounts of calcium carbonate remaining in the JBR underflow slurry. The gypsum solids settled very rapidly with no measurable differences in the free-fall characteristics between samples.

Figure 8 shows typical differences between solids formed when testing with lower sulfur coal (nominal 1.8 percent sulfur) and those formed with higher sulfur coal (nominal 3.2 percent sulfur). The crystals formed when cleaning the flue gas from



JBR underflow solids 11/29/78
1.8 percent sulfur coal



JBR underflow solids 4/13/79
3.2 percent sulfur coal

Figure 8. Comparison of solids produced with two SO_2 loading.

the lower sulfur coal were long rod-shaped crystals. Many were over 400 microns in length with length-to-diameter (L/D) ratios of from 10 to 20. The crystals produced with higher sulfur loadings were less than 100 microns with L/D ratios of from 2 to 5. This difference is consistent with what would be expected from operation with higher sulfate liquor loadings caused by the higher sulfur coal.

Operation with higher sulfur coal also increased the relative supersaturation of gypsum in the JBR although scaling conditions were not noted during the program. The maximum relative saturation measured, even during SO_2 spiking experiments (3000-3500 ppm SO_2 concentration), was only 1.23. This is well beneath the critical level of 1.3 which has been identified as the threshold for incipient scaling.[3]

Inspections at the conclusion of Phases I and II revealed little scale deposition in the JBR. There were some random patches of gypsum scale on various surfaces, but none of the depositions were threatening system performance and the scale thickness was less than 1/16 inch (2 millimeters). Since the scale buildup was minimal, infrequent routine cleaning might be necessary since the scale deposition will be a continuing phenomenon. The duration between system cleanings was not determined in the evaluation program, but it can be noted that nine months of operation were logged and no operating difficulties were experienced.

Gypsum Disposal. Throughout the program the gypsum was disposed of in a stack which is a disposal technique commonly used for gypsum produced in the phosphate fertilizer industry. Figures 9 and 10 show the stack during the initial fill period and in the final configuration. Chiyoda also tested the product for use in wallboard and Portland cement production. U.S. Gypsum and National Gypsum both made successful production runs of over 100 tons each with gypsum produced by the CT-121 prototype system. Laboratory tests also indicated that CT-121 gypsum could be used successfully in Portland cement. Details of the gypsum disposal testing will be the subject of a paper available with the handouts of this symposium.

EPA Performance Parameters. The four performance parameters employed by EPA to measure an FGD system's dependability are presented in Table 1. The overall figures include both the Phase 0 shakedown and the Phase III test period. Both of these periods involved some planned outages which penalized both the operability and utilization factors. However, during the EPRI program, all four factors were extremely good. There were only 22 hours of forced outages during the EPRI evaluation program. Of this, 21 hours were due to limestone feeder problems.

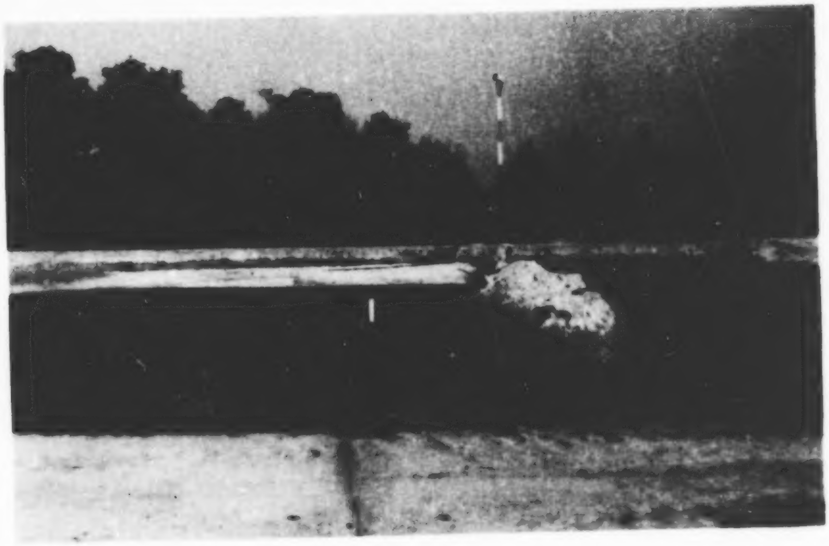


Figure 9. Stacking pond at start of Program 11/15/78.

Figure 10. Filled stack, end of Phase II 5/22/79.

Table 1

CT-121 VIABILITY PARAMETERS

	Viability Parameters (percent)			
	Availability[a]	Reliability[b]	Operability[c]	Utilization[d]
Chiyoda Shakedown Phase (Phase 0)	99.2	99.1	88.0	88.0
EPRI Evaluation Program (Phases I and II)	99.3	99.3	97.3	97.3
Extended Chiyoda Testing (Phase III)	99.5	99.1	58.6	58.6
Total Program Average	99.3	99.2	90.0	90.0

[a] Availability - Hours the FGD system is available for operation (whether operated or not), divided by the hours in the perid.

[b] Reliability - Hours the FGD system was operated divided by the hours the FGD system was called upon to operate.

[c] Operability - Hours the FGD was operated divided by the boiler operating hours in the period.
Total Program = 6552/7276

[d] Utilization Factor - Hours that the FGD system operated divided by total hours in the period.
Total Program = 6552/7276

When inspection time was added to the total downtime, the fraction of the period in which the FGD system operated was 97.3 percent as reflected in the utilization for Phases I and II.

These performance parameters indicate the CT-121 prototype performed with exceptional reliability during the evaluation program. These figures cannot be used to accurately predict the performance of a commercial system, but the evaluation program indicates that a properly designed CT-121 system could be expected to operate with a minimum of process or mechanical problems.

Mist Eliminator Performance. The mist eliminator performance during the program deserves special mention. The mist eliminator was composed of two banks vertical Chevron blades mounted in a horizontal run of duct downstream of the JBR. This mist eliminator was washed on an average of once a week for one minute with about 300 gpm 0.072 m^3/s of pond water. No signs of gypsum scaling or plugging were noted during the program.

This excellent performance is attributed to two major factors. First, the superficial gas velocity leaving the froth zone of the JBR was only about 2 ft/s resulting in most of the entrained slurry being separated from the flue gas in the interior of the JBR or in the JBR outlet gas chamber. Secondly, the slurry contained very little solid phase alkalinity ($CaCO_3$ or $CaSO_3$). Therefore, the dissolution of calcium solids and sorption of SO_2 on the mist eliminator blades which has caused scaling problems in many systems did not occur in the CT-121 prototype.

Overall System Controllability

The effective performance of the prototype system during the evaluation program was due to (1) the flexibility of the prototype to withstand process fluctuations, and (2) the controllability of the prototype system. Two key process control variables are monitored in the CT-121 system to ensure good performance: (1) JBR overflow pH and (2) JBR overflow solids concentration. pH was used as the primary control for SO_2 removal efficiency.

pH was continuously monitored with a dip-type sensor in the overflow weir, and limestone feed rate was manually adjusted based on this reading. A neoprene wiper was used to keep a stagnant film from building up around the probe. This instrument was checked daily and calibrated weekly.

pH fluctuations remained within ±0.2 units even after flue gas flow rate changes. Operation in the 3.0 to 4.5 pH range resulted in rapid limestone dissolution and good pH control.

The solids inventory in the JBR underflow was monitored by a nuclear density meter, and the gypsum discharge rate from the JBR was adjusted based on the solids concentration. Every four hours the operator checked the instrument by taking a slurry sample and measuring the volume of the settled solids. Although the solids content did vary somewhat from the set point, deviation from this set point did not cause scaling during the program. This was true even though the solids content was reduced significantly for a period of several hours during two different short-term tests.

It is also noteworthy to mention that locally hired operators were employed to actually run the process (2 operators per shift) and Chiyoda personnel were present only during the day shift for most of the program. The process operated with a minimum of problems or upsets using this approach to operator staffing.

SUMMARY

As a result of the independent evaluation program and related engineering activities, several CT-121 process design and operating features have been identified which may result in improved operability and reduced operating costs relative to existing lime/limestone systems:

- no large slurry recirculation pumps,
- no nozzles or screens,
- high limestone utilization,
- less dependence on limestone source and size on operation due to the low operating pH,
- low slurry entrainment in the gas enhancing mist eliminator performance,
- low scrubber profile which may lower capital costs and,
- the ability to operate successfully over a wide range of operating conditions with a minimum of scale deposition.

The concept of the JBR, therefore, represents a potentially attractive alternative to other currently available FGD technologies. The prototype at Scholz was successfully tested over a ten month period and was shown to operate reliably and efficiently under a variety of test conditions while treating flue gas from a coal-fired utility boiler.

REFERENCES

1. Randall E. Rush and Reed A. Edwards. Evaluation of Three 20 MW Prototype Flue Gas Desulfurization Processes. Final Report. Birmingham, AL: Southern Company Services, Inc. March 1978. EPRI No. FP-713-SY, EPRI Project No. RP536-1.

2. H. Idemura, T. Kanai, and H. Yanagioka. "Jet Bubbling Flue Gas Desulfurization." Chemical Engineering Progress. February 1978. P. 46-50.

3. D. M. Ottmers, Jr., et al. A Theoretical and Experimental Study of the Lime/Limestone Wet Scrubbing Process. Austin, TX: Radian Corp. 1974. EPA 650/2-75-006, EPA Contract No. 68-02-0023.

Forced Oxidation of Limestone Scrubber Sludge at TVA's Widows Creek Unit 8 Steam Plant

by

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ABSTRACT

Tests on one module (140 MW) have been carried out to demonstrate the feasibility of forced oxidation of limestone scrubber sludge to gypsum as a viable technique for ultimate disposal of these waste materials. Both one-tank and two-tank oxidation experiments were studied with data indicating the two-tank runs more closely met test objectives. Equations to predict oxidation were developed and expressed as a function of mass transfer and chemical kinetics. Air stoichiometries of between 1.75 and 2.0 lb atoms O/lb mole SO_2 absorbed will consistently produce oxidation of ~95%.

As a result of the Forced Oxidation Test Program, this method is being given consideration as one of the alternative methods of scrubber sludge disposal for Widows Creek units 7 and 8. Additionally, Paradise Steam Plant units 1 and 2 scrubber trains are being designed with a forced oxidation option to produce a sulfate waste product.

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FORCED OXIDATION OF LIMESTONE SCRUBBER SLUDGE AT TVA'S WIDOWS CREEK UNIT 8 STEAM PLANT

INTRODUCTION

TVA uses, or plans to use, limestone wet flue gas scrubbing as the method of reducing SO_2 emissions at two of its twelve coal-fired steam-electric generating plants--Widows Creek units 7 and 8 and Paradise units 1 and 2. At the remaining ten coal-fired plants, Widows Creek units 1-6, and Paradise unit 3, TVA burns either low- or medium-sulfur coal, or washed coal to achieve the required SO_2 emission limitations.

Forced oxidation will be utilized in the disposal of the sludge at the Paradise plant. The scrubbers are scheduled to become operational by September 1982. Forced oxidation is presently being compared and evaluated with other sludge disposal methods at Widows Creek. The total life-cycle costs of forced oxidation sludge disposal will be compared with total life-cycle costs of the alternatives of raw ponding, mixing the sludge with dry fly ash, and mixing the sludge with dry fly ash plus additives.

The future role of forced oxidation depends primarily on the following:

1. Technical feasibility of the process.
2. The total life-cycle costs of forced oxidation as compared with other disposal methods.
3. The final requirements of the Resource Conservation and Recovery Act (RCRA) on disposing of scrubber sludge.

BACKGROUND

For the last several years, TVA has been involved in an intensive research and development project which was initiated to make a thorough and complete assessment of its first full-scale scrubber system at Widows Creek unit 8, located near Stevenson, Alabama. The research and development effort consisted of six tasks which were designed to evaluate the scrubber system. This paper will report results of the forced oxidation experiments at Widows Creek unit 8.

The wet limestone scrubber system, designed and constructed by TVA, treats flue gas from a 550-MW Combustion Engineering (CE) tangentially coal-fired boiler. The flue gas desulfurization (FGD) system consists of four parallel scrubber trains, each capable of scrubbing 25 percent of the flue gas. Only one of the four scrubber trains, train D, was used for the forced oxidation demonstration experiments. Assistance in identifying the design criteria for the test program of the forced oxidation demonstration was obtained from the studies performed at the Shawnee Test Facility.

TVA contracted with CE to install forced oxidation equipment on the FGD system at the Widows Creek Steam Plant to demonstrate that forced oxidation of FGD wastes is possible at this location as a processing scheme for waste disposal. The forced oxidation demonstration program began on April 2, 1979, and continued until November 15, 1979. CE had the responsibility for the initial operational phase through June 30, after which Radian Corporation assumed the operational responsibility through November 15, 1979.

A flow schematic of Widows Creek Unit 8 Wet Limestone Scrubber System is shown in Figure 1. The pressurized scrubber system consists of four identical trains located downstream of low efficiency (approximately 50 percent) electrostatic precipitators. The principal components in each train A, B, C, and D include a boiler I.D. fan, venturi scrubber, grid-type spray tower absorber, Chevron-type entrainment separator, indirect steam reheat system, venturi slurry recirculation system, and absorber slurry recirculation system. Each module is capable of treating approximately 25 percent of the boiler flue gas at full load.

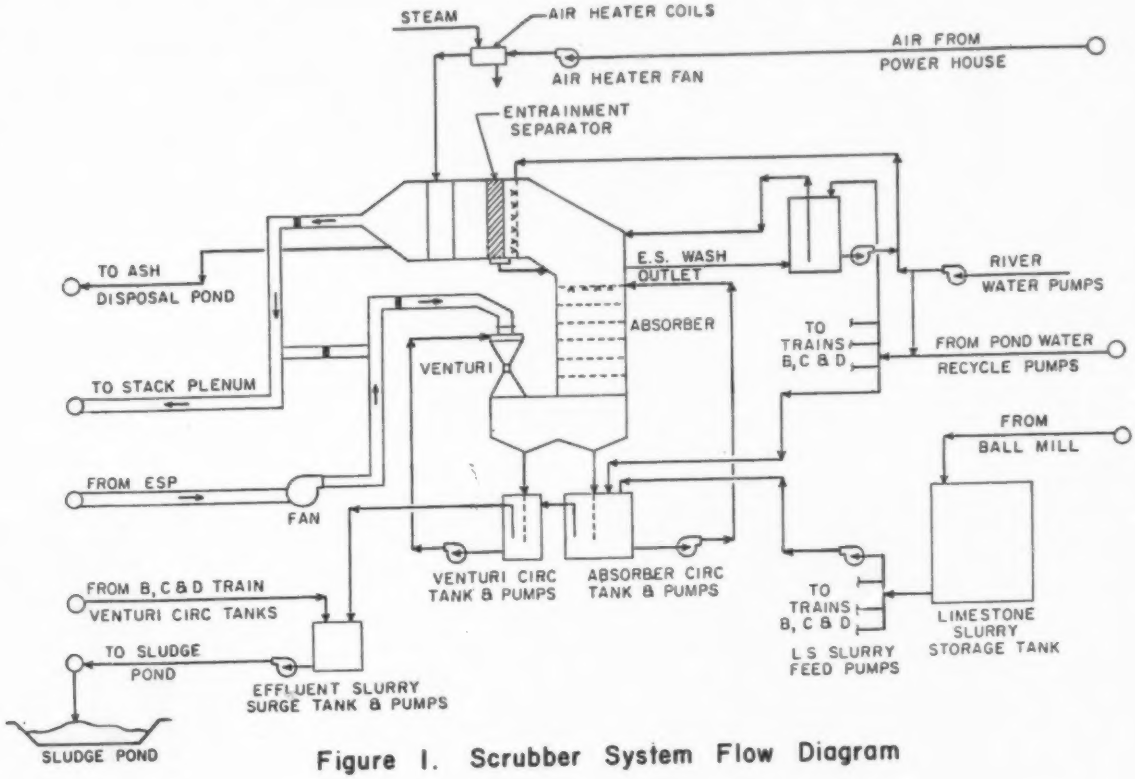

STEAM
AIR HEATER COILS
AIR FROM
POWER HOUSE
AIR HEATER FAN
ENTRAINMENT
SEPARATOR
E.S. WASH
OUTLET
RIVER
WATER PUMPS
TO ASH
DISPOSAL POND
ABSORBER
TO
TRAINS
B, C & D
FROM POND WATER
RECYCLE PUMPS
VENTURI
TO STACK PLENUM
FROM
BALL MILL
FROM ESP
FAN
VENTURI CIRC
TANK & PUMPS
ABSORBER CIRC
TANK & PUMPS
FROM B, C & D TRAIN
VENTURI CIRC TANKS
TO
TRAINS
B, C & D
LIMESTONE
SLURRY
STORAGE TANK
TO SLUDGE
POND
L S SLURRY
FEED PUMPS
EFFLUENT SLURRY
SURGE TANK & PUMPS
SLUDGE POND

Figure 1. Scrubber System Flow Diagram

The waste slurry produced by the FGD system currently is stored in a 110-acre pond. Disposal of this slurry represents a major problem in continued, long-term operation of the scrubber unit. It was decided to evaluate and demonstrate the forced oxidation method for treating the sludge to decrease the effective volume required for disposal or to improve the stability of material in the disposal area. During the demonstration program, a forced oxidation system was installed on scrubber train D. Approximately 10 percent (4,370 lb/hr of solids) of the oxidized scrubber bleed stream was processed through a 2-stage dewatering system consisting of a thickener and rotary drum vacuum filter. A flow schematic of the demonstration unit is shown in Figure 2. Initially, the filter cake from the vacuum filter was reslurried and disposed of in the pond. After the test objectives had been achieved and conditions determined for producing a cake of consistent quality, arrangements were made to initiate a landfill disposal project for long-term monitoring of the final product. The gypsum produced at Widows Creek is unusual in that 30 percent of the solids is fly ash. The presence of this ash may have as yet undetermined effects on the long-term stability of the final disposal material.

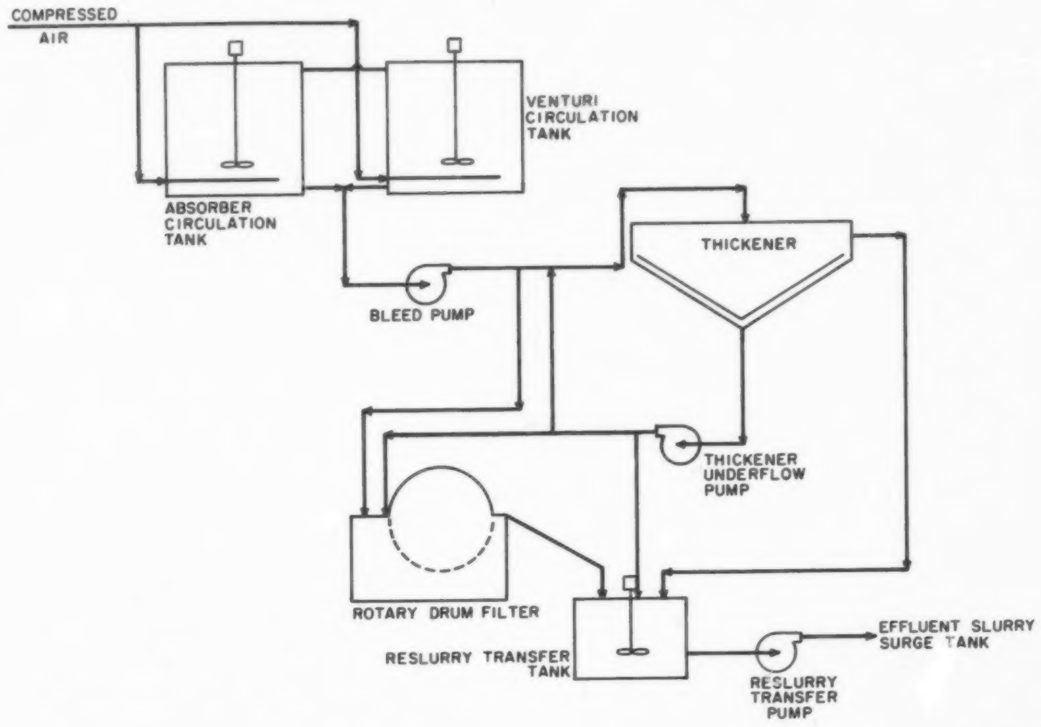

COMPRESSED AIR
VENTURI CIRCULATION TANK
ABSORBER CIRCULATION TANK
THICKENER
BLEED PUMP
THICKENER UNDERFLOW PUMP
ROTARY DRUM FILTER
RESLURRY TRANSFER TANK
EFFLUENT SLURRY SURGE TANK
RESLURRY TRANSFER PUMP

Figure 2. Forced Oxidation Dewatering Equipment

Oxidizing air was introduced to both the venturi and absorber tanks by means of a circular sparge ring, located just beneath the agitator impellers as shown in Figures 3 and 4. Air was discharged through thirty-six 1-1/4-inch holes on the outside of the sparge ring (Figure 5).

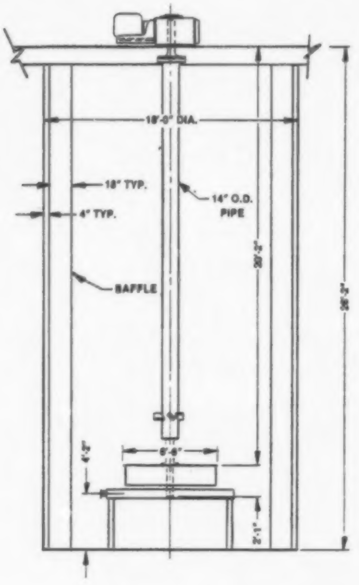

18'-0" DIA.
18" TYP.
4" TYP.
14" O.D. PIPE
BAFFLE
6'-6"
4'-3"
2'-1"

Figure 3. Venturi Tank Agitator and Sparge Ring

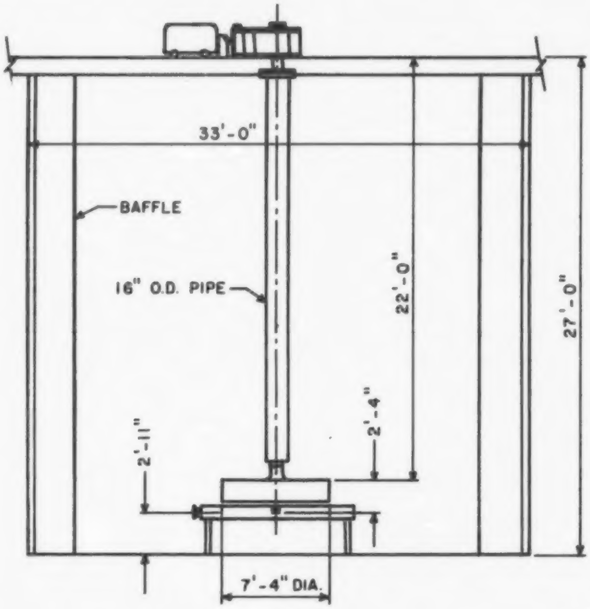

33'-0"
BAFFLE
16" O.D. PIPE
22'-0"
27'-0"
2'-4"
2'-11"
7'-4" DIA.

Figure 4. Absorber Tank Agitator and Sparge Ring

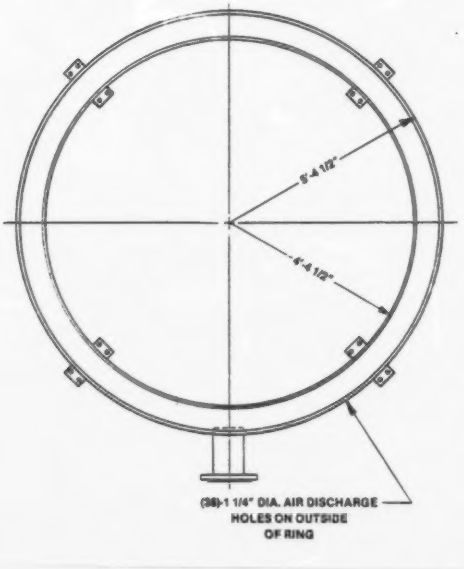

5'-4 1/2"
4'-4 1/2"
(36)-1 1/4" DIA. AIR DISCHARGE HOLES ON OUTSIDE OF RING

Figure 5. Absorber Tank Sparge Ring

Objectives and Goals

The general objectives of the test program were as follows:

- Demonstration of forced oxidation as a viable FGD waste disposal option.
- Acquisition of data applicable to the design of a forced oxidation system for the Widows Creek units 7 and 8 FGD systems.

Three specifically quantitative goals of the demonstration program were as follows:

- Attainment of greater than 95 mole percent conversion of calcium sulfite to sulfate.
- Production of a waste product capable of being dewatered to greater than 80 wt percent solids with a vacuum filter.
- SO_2 removal efficiency of 88 percent or greater.

Originally, the test plan for the forced oxidation experiments was divided into three separate test blocks. One test block (experiment B) was designed to verify the hypothesis that the air stoichoimetry required for oxidation in the venturi effluent hold tank only would be significantly less than in the absorber tank. Results of this type have been observed at the Shawnee test facility. However, such was not the case due presumably to excessive carryover of venturi-loop liquors into the absorber loop. This carryover appears to be a function of boiler load (gas velocity) and results in either a levelized pH in both tanks or inversions such that the absorber tank actually had at times a lower pH than the venturi tank. Oxygen stoichiometry requirements are closely related to pH.

A second test block (experiment C) involved oxidation in both the venturi and absorber hold tanks simultaneously to determine if such dual tank oxidation could be accomplished at a lower oxygen stoichiometry than single tank oxidation. A third series of runs (experiment A) with oxidation in the absorber tank only was cancelled because of the difficulties experienced with one tank (venturi) oxidation as described in experiment B above.

Test Description and Results

Operating conditions were varied to meet the three goals of primary interest. Oxidation, dewatering, and SO_2 removal were each studied with minimal interference from the other parameters. To determine the minimum amount of air necessary to achieve 95-percent oxidation, an initial air rate was chosen which was known to give about 99-percent oxidation. The air rate was then reduced stepwise until oxidation was consistently above 95 percent. After tests at several air rates which gave between 91 and 99 percent oxidation, it was possible graphically to determine the minimum air stoichiometry.

The air stoichiometry was then held constant at this value while the dewatering train was tested. This involved determination of the appropriate combination of feed rate, filter area, drum speed, and filter pressure that produced a product containing a minimum of 80 percent solids.

Since the FGD unit is a commercial system, it was difficult to vary a large number of operating variables to increase SO_2 removal. In addition, TVA was simultaneously involved in a comprehensive test program aimed at improving the SO_2 removal of the Widows Creek FGD system. It was felt that the results of this companion program could be applied to a forced oxidized system as well as the existing system. Consequently, only two variables were examined in regard to SO_2 removal: Ca/S ratio and the effect of additional packing in the absorber. Testing of these two parameters was sufficient to produce the desired 88-percent SO_2 removal. During the SO_2 removal tests, the forced oxidation unit was operated to further refine the operating parameters. The equipment was operated in combination to demonstrate that the FGD unit and oxidation unit could produce oxidized sludge of the desired quality while meeting the SO_2 removal requirements.

Operating parameters were set on the desired conditions and a period of 8 hours was used to allow steady state operation before sampling. Usually only one sample per sample point per day was drawn and analyzed. An additional 12 to 14 hours of operation was required to verify steady state conditions and to obtain a second set of chemical data. A summary of sample data and analytical determinations is given in Table 1.

Table 1. Summary of Analyses

Sample Phase	Bleed Stream	Thickener Underflow	Thickener Overflow	Vacuum Filter Drum	Absorber Bottoms
Liquid	$\left.\begin{array}{l}SO_3^=\\SO_4^=\\NO_3^-\\Cl^-\\CO_3^=\end{array}\right\}$ Anions $\left.\begin{array}{l}Ca^{++}\\Mg^{++}\\Na^+\\K^+\\NH_4^+\end{array}\right\}$ Cations				$\left.\begin{array}{l}SO_3^=\\SO_4^=\\NO_3^-\\Cl^-\\CO_3^=\end{array}\right\}$ Anions $\left.\begin{array}{l}Ca^{++}\\Mg^{++}\\Na^+\\K^+\\NH_4^+\end{array}\right\}$ Cations
Solid	$\left.\begin{array}{l}SO_3^=\\SO_4^=\\CO_3^=\end{array}\right\}$ Anions $\left.\begin{array}{l}Ca^{++}\\Mg^{++}\end{array}\right\}$ Cations Acid Insoluble Quantity Wt % Solids	$\left.\begin{array}{l}SO_3^=\\SO_4^=\\CO_3^=\end{array}\right\}$ Anions $\left.\begin{array}{l}Ca^{++}\\Mg^{++}\end{array}\right\}$ Cations Acid Insoluble Quantity Particle Size Distribution Wt % Solids	% Solids	% Solids Thickness on Drums	$\left.\begin{array}{l}SO_3^=\\SO_4^=\\CO_3^=\end{array}\right\}$ Anions $\left.\begin{array}{l}Ca^{++}\\Mg^{++}\end{array}\right\}$ Cations Acid Insoluble Quantity Wt % Solids
Slurry	pH Temp Settling Rate Test	pH Temp Filter Leaf Test	pH Temp		pH Temp

The following methods were employed to determine quantitatively cations and anions of both solid and liquid samples.

Cation Analyses

Analyses for calcium, magnesium, sodium, and potassium were done by atomic absorption spectrophotometry.

Anion Analyses

The analyses for the anions required a number of different analytical methods. These methods included:

- Ion chromatography for sulfate and chlorine determinations.
- Iodometric "back" titration with iodine for sulfite determinations.
- Nondispersive infrared (NDIR) analysis for CO_2 determinations.

A representative set of operational and chemical data collected during the demonstration is shown in Table 2.

TABLE 2. BLEED STREAM ANALYSIS AND TEST RESULTS

Liquid Phase

					Milligrams/liter					
Date	Time	Test Condition	Temp °C	pH	$SO_3^{=}$	$SO_4^{=}$	Cl^-	$CO_3^{=}$	Ca^{++}	Mg^{++}
12/20/79	0745	C5	41	5.7	19	1278	1021	144	770	163

Solid Phase

				Milligrams/gram solid					
Date	Time	Test Condition	Venturi pH	Ca^{++}	$SO_3^{=}$	$SO_4^{=}$	$CO_3^{=}$	Mg^{++}	Acid insoluble
12/20/79	0745	C5	5.7	177	3	303	91	3	279

Test Results

Relative saturation $CaSO_3 \cdot \frac{1}{2}H_2O$	Relative saturation $CaSO_4 \cdot 2H_2O$	% Residual electroneutrality	Comments	Ca/SO_x ratio $\frac{\text{lb-mol}}{\text{lb-mol}}$
0.2	0.7	-3.6	Steady state	1.4

$CaCO_3$ added $\frac{\text{lb-mol}}{\text{min}}$	SO_2 absorbed $\frac{\text{lb-mol}}{\text{min}}$	Oxygen added $\frac{\text{lb-atoms}}{\text{min}}$	O/SO_2 ratio $\frac{\text{lb-atoms}}{\text{lb-mol}}$	Percent oxidation	Vacuum filter cake % solids
2.2	1.3	2.8	2.3	99	86.8

CHEMISTRY OF FORCED OXIDATION

Proposed Reaction Mechanisms

Oxidation may best be explained in terms of mass transfer and chemical kinetics. The oxidation of sulfite to sulfate can be affected by several chemical phenomena including gas absorption, reaction kinetics, dissolution of solids and precipitation of solids. The overall reaction,

$$CaSO_3 + \tfrac{1}{2}O_2 \rightarrow CaSO_4 \qquad (1)$$

can be broken down into several steps, each a necessary link of the reaction pathway and, as such, each a potential rate limiting step. These are shown in Table 3.

TABLE 3. REACTION STEPS AND PHENOMENA INVOLVED IN SULFITE OXIDATION

Reaction Step	Phenomena Involved
Dissolution of reactants	Gas absorption, solids dissolution
Reaction	Kinetics
Precipitation of products	Solids precipitation

First, it is necessary for the reactants to be in the liquid phase. This involves absorption of O_2 by the scrubbing liquor and either dissolution of solid calcium sulfite or absorption of SO_2 as indicated by reactions (2) through (5).

$$O_2(g) \rightarrow O_2\ (aq) \qquad (2)$$

$$SO_2(g) + H_2O \rightarrow HSO_3^-(aq) + H^+(aq) \qquad (3)$$

$$CaSO_3 \cdot \tfrac{1}{2}H_2O(s) \rightarrow Ca^{++} + SO_3^= + \tfrac{1}{2}H_2O \qquad (4)$$

$$SO_3^= + H_2O \rightarrow HSO_3^- + OH^- \qquad (5)$$

The oxidation reaction then takes place between the dissolved reactants according to reaction (6).

$$HSO_3^-(aq) + \tfrac{1}{2}O_2(aq) \rightarrow SO_4^=(aq) + H^+(aq) \qquad (6)$$

Finally, the reaction product, gypsum, is removed from the liquid phase by precipitation.

$$SO_4^=(aq) + Ca^{++}(aq) + 2H_2O \rightarrow CaSO_4 \cdot 2H_2O(s) \qquad (7)$$

The entire sequence is illustrated schematically in Figure 6 and each of these steps is discussed in detail in the following paragraphs.

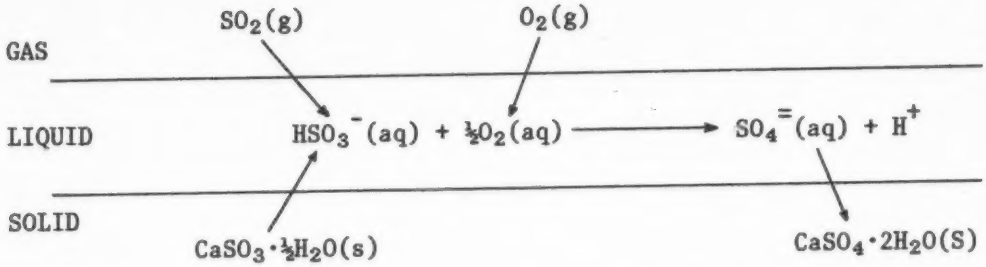

GAS
LIQUID
SOLID
$SO_2(g)$
$O_2(g)$
$HSO_3^-(aq) + \frac{1}{2}O_2(aq) \longrightarrow SO_4^=(aq) + H^+$
$CaSO_3 \cdot \frac{1}{2}H_2O(s)$
$CaSO_4 \cdot 2H_2O(S)$

Figure 6. Mass Transfer and Reaction Sequence

Dissolution of Reactants

Bisulfite, HSO_3^-, appears to be the reactive form of the dissolved $CaSO_3$, hence subsequent discussion will refer to bisulfite rather than sulfite which is also a dissolved species.

Bisulfite can be dissolved in the liquid phase by two mechanisms, absorption of SO_2 from the flue gas and dissolution of solid calcium sulfite.

In forced oxidation processes where oxidation takes place outside the scrubber circulation loop, dissolution of calcium sulfite solids is clearly the predominant form of bisulfite generation. Since the bisulfite is being removed by oxidation, the sulfite relative saturation is expected to be low (<1) and calcium sulfite dissolves.

The Widows Creek demonstration involved oxidation within the scrubber loop. In this case, bisulfite ions will be supplied by the absorber, and dissolution of solid sulfite will not be of any importance unless sulfite solids are precipitating in some part of the system. During C-series testing, air was sparged into both hold tanks. Figure 7 is a plot of the relative saturation of calcium sulfite versus percent oxidation in the solid-phase for the B and C series tests.

$$RS_{CaSO_3 \cdot \frac{1}{2}H_2O} = \frac{a_{Ca^{++}} \cdot a_{SO_3^=} \cdot a_{H_2O}^{\frac{1}{2}}}{K_{sp_{CaSO_3 \cdot \frac{1}{2}H_2O}}} \qquad (8)$$

where

a = activity of subscripted component

K_{sp} = solubility product of the precipitation reaction (temperature dependent)

Note that for C-series tests (two tank oxidation) the relative saturation was substantially greater than 1 at most oxidation levels, indicating that bisulfite ions absorbed in the scrubber were oxidized without first precipitating as $CaSO_3 \cdot \frac{1}{2}H_2O$. This means that at intermediate oxidation levels, oxidation competes with precipitation for HSO_3^- ions, and dissolution of solid sulfite is not part of the reaction mechanism. The low relative saturations seen at extremely high oxidation levels are most likely due to sulfite ion depletion, caused by sulfate precipitation and not any dissolution mechanism.

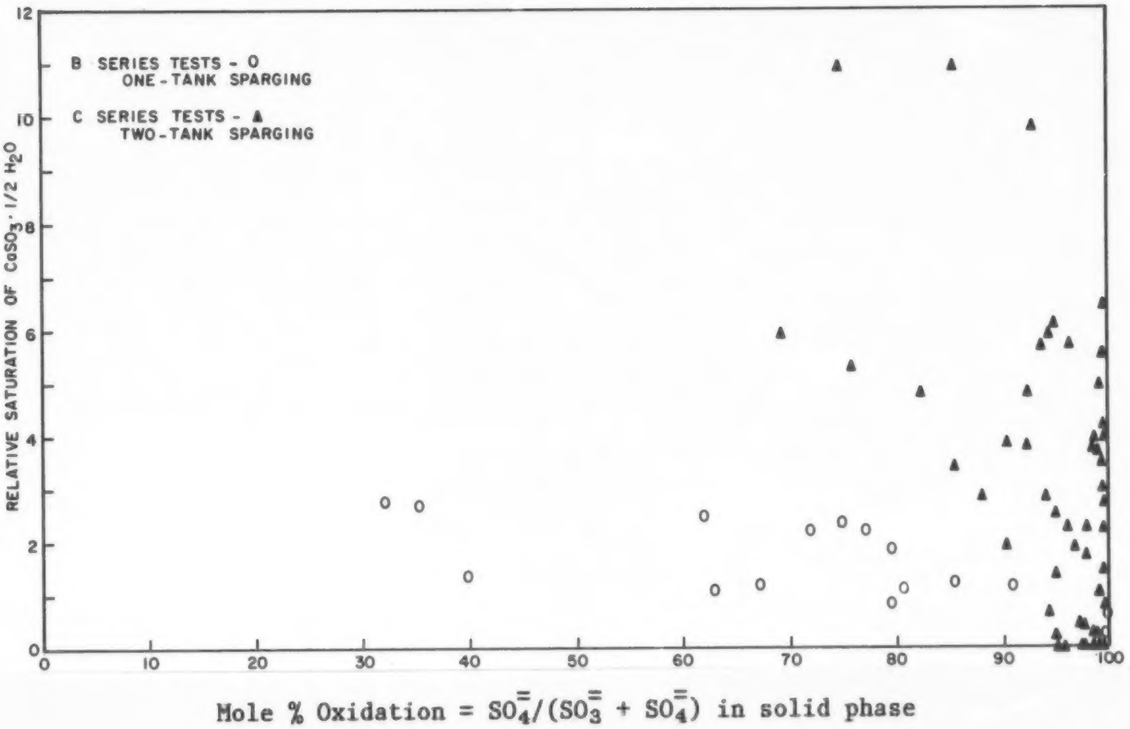


Figure 7. Relative Saturation of $CaSO_3 \cdot \frac{1}{2}H_2O$ vs Oxidation in Venturi Hold Tank - One- and Two-Tank Air Sparging

Relative saturation data for B-series tests (one tank oxidation) indicate that dissolution may have been a more important factor in these tests. With air introduced only into the venturi circulation tank, $CaSO_3 \cdot \frac{1}{2}H_2O$ precipitation would have been expected in the absorber circulation tank, necessitating dissolution in the venturi tank. Also, since significant backmixing occurred between the two tanks, the dissolution step has been made more pronounced as a source of calcium bisulfite than for a true two-stage system with single-tank oxidation.

In summary, this analysis indicates if the absorber tank is sparged, then the source of bisulfite is SO_2 absorption and solids dissolution is not necessary. Therefore, solids dissolution can be eliminated as a potential rate limiting step in the C-series testing described in this report. However, this is not true for the B-series testing performed earlier. When just the venturi tank is sparged, solids dissolution is a necessary link in the reaction pathway since the source of bisulfite resulting from SO_2 absorption is insignificant.

Another step of the reaction pathway involving reactant dissolution is oxygen absorption. This is primarily a mass-transfer process, affected by both chemical and mechanical factors. The system chemistry affects two parameters that are important to mass transfer: oxygen solubility and the diffusivity of dissolved oxygen. Oxygen solubility decreases with increases in both temperature and ionic strength. The diffusivity of dissolved oxygen will increase with increasing temperature and decrease with increasing ionic strength. Neither temperature nor ionic strength varied substantially during demonstration testing.

pH also has an important effect on the oxidation reaction in that oxidation of sulfite causes the pH to rise when in the presence of excess limestone. This increase in pH will rapidly quench the reaction unless some method of replenishing H^+ ions is available. At Widows Creek, the pH level for oxidation purposes was maintained by SO_2 absorption.

Under normal operating conditions at Widows Creek, with the pH at 5 to 6, the sulfite species are 40 to 70 percent in the bisulfite form. This is sufficient for forced oxidation. The pH's of the two hold tanks rarely differed by more than 0.5, but if the backmixing problem were eliminated, the pH in the absorber would be higher. Since bisulfite availability is not rate-limiting, the effect this would have on a force-oxidized system may be small unless the pH was over 6.5.

The data indicate that during the tests, pH values were in the proper range to provide high concentrations of the bisulfite ion. Therefore, bisulfite ion availibility can be eliminated as a limiting step in the overall reaction sequence. The potential rate-limiting steps are thus reduced to reaction kinetics, product removal, or O_2 transfer.

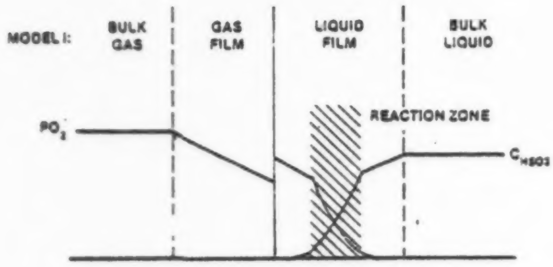
Reaction Kinetics

Thermodynamics indicate that the oxidation (equation 5) reaction is essentially irreversible, with an equilibrium constant (25°C) on the order of 10^{40}.[1] Therefore, we need only consider the kinetics of the forward reaction.

Most forced-oxidation research has indicated that the reaction is relatively fast. If this is true, then the site of the reaction is limited to the gas/liquid interface. This indicates that mass transfer to the film between the gas and the bulk liquid will be the limiting factor, and not reaction kinetics.

Using film theory terminology, the phase interface and reaction zone of the oxidation reaction can be shown by two models in Figure 8. Model I illustrates a fast second order rate reaction where the diffusion rate of O_2 and HSO_3^- in the liquid film are not limiting the overall reaction and the reaction is fast enough so that the reaction zone remains totally within the liquid film.

Model II shows an instantaneous reaction first order rate expression where the concentration of HSO_3 is relatively high and the reaction plane is moved to the gas-liquid interface. The overall rate will be limited by the diffusion of O_2 through the gas film. The film models shown in the figure result in mathematical models for the reaction rate that are essentially the same as those for mass transfer without reaction, except for an "enhancement" term. Reaction in the film tends to thin the film, and the enhancement term accounts for the corresponding increase in mass transfer.


MODEL I:
BULK GAS
GAS FILM
LIQUID FILM
BULK LIQUID
PO_2
REACTION ZONE
C_{HSO_3}

Reaction only in film, general fast-irreversible reaction.

R = reaction rate = $K_g$eaPO₂

$K_g e$ = $\left[\frac{1}{K_{O_2g}} + \frac{H_{O_2}}{K_{O_2l}E}\right]^{-1}$

K_{O_2g}, K_{O_2l} = Gas- and liquid-side oxygen mass-transfer coefficients

H_{O_2} = Henry's Law constant for oxygen

E = Complex function of $C_{HSO_3^-}$, C_{O_2}

a = Interfacial area per unit volume

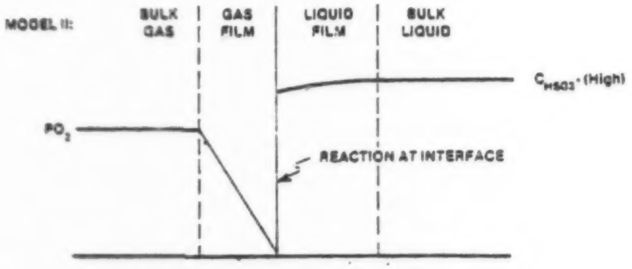
PO_2 = Gas-side partial pressure O_2

For an infinitely fast reaction, the expressions become:

R = $K_g a\ E\ PO_2$

K_g = $\left[\frac{1}{K_{O_2g}} + \frac{H_{O_2}}{K_{O_2l}}\right]^{-1}$

E = $\left[1 + (\text{constant}) \times \frac{C_{HSO_3^-}}{PO_2}\right]$


MODEL II:
BULK GAS
GAS FILM
LIQUID FILM
BULK LIQUID
C_{HSO_3} (High)
PO_2
REACTION AT INTERFACE
02-6110-1

Reaction only in film, high HSO_3^- concentration, general fast-irreversible reaction.

$R = K_g a\ PO_2$

$K_g = \left[\frac{1}{K_{O_2g}} + \frac{\text{constant}}{\sqrt{C_{HSO_3^-}}}\right]^{-1}$

For an infinitely fast reaction, the expressions become:

$R = K_{O_2g}\ a\ PO_2$

Figure 8. Models for Film Reactions[2]

The film reaction models suggest that mass transfer of O_2 is the limiting step rather than kinetic limitations. This is evidenced by the presence of the oxygen mass transfer coefficient in the expressions for reaction rate constant. Therefore, in terms of the overall reaction rate, both sulfite and reaction kinetics are not rate-limiting steps. Sulfate precipitation rate does not affect the kinetics and, as will be shown in subsequent discussion, will be a function of the reaction rate. The most probable rate-limiting step in the overall reaction sequence is then oxygen transfer. However, data are needed to confirm this and to validate one of the models shown in Figure 11.

Removal of Products

An important aspect of forced oxidation is its effect on the relative saturation of $CaSO_4$. The relative saturation for sulfate (RS) is defined similarly to that for sulfite.

$$RS_{CaSO_4 \cdot 2H_2O} = \frac{a_{Ca^{++}} \cdot a_{SO_4^=} \cdot a^2_{H_2O}}{K_{sp_{CaSO_4 \cdot 2H_2O}}} \qquad (9)$$

where

a = activity of subscripted component

K_{sp} = solubility product of the precipitation reaction (temperature dependent)

RS of calcium sulfate is important because it can affect scaling of the scrubber. Generally, for $1 < RS < 1.3$, crystal growth will occur on existing gypsum crystals. At $RS > 1.3$ gypsum nucleation can occur which can result in crystal growth or scaling on the scrubber internals. This can eventually require a shutdown to remove the scale.

While forced oxidation increases the total amount of $SO_4^=$ or $CaSO_4 \cdot 2H_2O$ present, the $RS_{CaSO_4 \cdot 2H_2O}$ at Widows Creek was not increased and may have been decreased. This is attributed to the higher sulfate concentrations resulting in an increase in gypsum solids surface area which will enhance the gypsum precipitation rate.

After the bisulfite is oxidized, it is necessary to remove the sulfate product from solution. This is accomplished via precipitation of the sulfate as solid gypsum, $CaSO_4 \cdot 2H_2O$. Gypsum precipitation has been studied extensively with respect to flue gas desulfurization systems. The driving force for precipitation is relative saturation, and three regimes can be considered as shown in Table 4.

TABLE 4. GENERAL EFFECT OF RELATIVE STATURATION OF GYPSUM PRECIPITATION

Relative Saturation	Effect
<1	Dissolution of gypsum solids
1.0 - 1.3	Precipitation on existing solids
>1.3	Solids nucleation

A simple rate expression for gypsum precipitation can be written as

$$r = KafC\ (RS-1) \tag{10}$$

where

r = precipitation rate (gram/liter-sec)

K = temperature dependent constant (gram/cm^2-sec-unit driving force)

a = crystal interfacial area per gram of precipitation solid (cm^2/gram)

f = weight fraction of the precipitation species in the solid phase

C = total solids concentration in the slurry (grams/liter)

RS = relative saturation of calcium sulfate in the liquid phase

The precipitation rate of $CaSO_4$ is a function of both total crystal surface area and relative saturation. At steady state the precipitation rate must equal the oxidation rate. From an operating standpoint, it is desirable for the relative saturation to be kept below 1.3 to avoid scaling of the scrubber internals. In a design situation, the maximum relative saturation can be limited by designing an appropriately sized reaction vessel. A large tank volume will give an equivalent precipitation rate at a lower relative saturation. However, in a retrofit situation, as exists at Widows Creek, the tank volume is fixed, and either crystal surface area or relative saturation will vary in order to provide the necessary precipitation rate.

The relative saturation of gypsum in the venturi hold tank was approximately 1 during most of the forced-oxidation testing at Widows Creek. This indicates that a relatively small driving force (RS-1) was sufficient to keep the crystallization rate equal to the oxidation rate. This result might have been anticipated because the available gypsum crystal surface area (afC) is high. The system may have benefited from the significant increase in agitation over nonforce-oxidized systems. The increased agitation may increase available crystal surface area by shearing existing crystals into smaller ones. Since a relatively small driving force was sufficient to keep the precipitation rate to equal the oxidation rate, this indicates that precipitation was not a rate-limiting step.

Oxygen Transfer

In a process consisting of several steps, the rate of the limiting step is essentially equal to that of the overall process. In this case, if the oxygen transfer rate is assumed limiting and can be modeled, then the rate of the overall process will be known. Furthermore, this analysis can determine which are the significant variables affecting this critical path. This information will allow accurate decisions to be made with respect to improving the sulfite oxidation step and perhaps yield data useful in applying forced oxidation to limestone scrubbers of different configurations. The analysis approach and results are presented in the following section.

Data Analysis

The preceding discussion of chemistry and mass-transfer theories suggests several variables that may have an effect on oxidation rates, since the reaction is apparently mass-transfer limited, and the variables that most directly affect mass-transfer should be of primary importance. These would include the partial pressure of oxygen, the concentration of HSO_3^-, and mass-transfer coefficients. Also, pH, temperature, and ionic strength could be expected to affect the oxidation rate because of effects on oxygen solubility and sulfite concentration.

Table 5 presents some of the variables calculated in order to model the forced oxidation process. The oxidation rate (R) calculation and the net air stoichiometry (AS) calculation account for a 32 percent "baseline" or "natural" oxidation that is not due to air sparging. The gross air stoichiometry (FAS) calculation implicitly assumes 100 percent forced oxidation and no baseline oxidation. Consequently, AS will be a somewhat higher number than FAS. FAS is the air stoichiometry usually presented in the literature when forced oxidation is discussed.

Two general types of models were examined, mass transfer and overall oxidation rate. The first type involved correlating measured and calculated variables with the corresponding calculated mass transfer coefficient ($K_g a$). The second type involved correlating measured and calculated variables with the calculated oxidation rate (R).

Several data fits were made based on these models, in order to determine the most accurate representation of the Widows Creek data. The results of the correlations of the data obtained during the demonstration are inadequate to model the rate of the oxidation process. This is not surprising when the nature of the data is considered.

Attempts to model the forced-oxidation system at Widows Creek were complicated by several problems. Modeling the reaction was not a program objective and, therefore, no attempt was made during demonstration testing to vary important parameters systematically over a broad range. Also, there was no reliable means of monitoring continuously the flue gas flow rate due to high grain loadings which plugged probes. Rather this was estimated by correlating fan amps

Table 5. Variables and Units - Oxidation Modeling

	Variable	Units	Source
R	Oxidation Rate ("Forced")	$\frac{\text{lb-mole S}}{\text{min}}$	$\left[2.64 \times 10^{-9} \times \text{Flue Gas (scfm)} \times \Delta SO_2\text{(ppm)} \times \left(\frac{Ox - 32}{100}\right)\right]$
Ox	Net Solid-Phase	mole %	$\left[\frac{\text{moles } SO_4^{=}}{\text{moles } SO_3^{=} + \text{moles } SO_4^{=}} \times 100\right]$ in solid-phase analysis
$K_g a$	Mass-transfer Coefficient	$\frac{\text{lb-mole } O_2}{\text{min-f}^3\text{-(psi driving force)}}$	$\left[\frac{\text{Air} \times (0.791)\text{(scfm)}}{\text{(tank volume f}^3) \times \text{Avg Pressure (psi)}} \times \ln\left(\frac{AS}{(AS-1)}\right)\right]$
AS	Net Air Stoichiometry	$\frac{\text{lb-atom O in}}{\text{lb-mole S force-oxidized}}$	$\left[\frac{(\text{Air} \times 0.209)\text{ scfm} \times 2\text{ atom/mole } O_2}{379\text{ f}^3\text{/lb-mole} \times R\text{ lb-mole S/min}}\right]$
FAS	Gross Air Stoichiometry	$\frac{\text{lb-atom O in}}{\text{lb-mole S removed}}$	$\left[\frac{(\text{Air} \times 0.209)\text{ scfm} \times 2\text{ atom/mole } O_2}{\text{Flue Gas (scfm)} \times \Delta SO_2\text{(ppm)} \times 10^{-6}}\right]$
PO_2	Partial Pressure O_2 (log mean)	psi	log mean (PO_2 in, PO_2 out) $= \left[\frac{PO_2\text{ in} - PO_2\text{ out}}{\ln \frac{PO_2\text{ in}}{PO_2\text{ out}}}\right]$
PO_2 in	Partial Pressure O_2 (bottom of tank)	psi	0.209(14.7 + hydrostatic head) = [5.31 psi]
PO_2 out	Partial Pressure O_2 (top of tank)	psi	$\left[(0.209 \times 14.7) \times \left(\frac{AS-1}{AS - 0.209}\right)\right]$ from mass balance

against periodic velocity traverses. This method is approximate at best, and the problem was compounded by frequent shutdowns of the unit 8 boiler. When the unit was derated to 300 MW, it became possible to operate with one of the four scrubber trains off-line by biasing the gas flow to the other trains. During this period, the ratio of fan amps to flue gas rate in D-train was radically different when A-train or B-train was off-line, versus when C-train was off-line. Flue gas flow rate measurements are used in SO_2 removal rate calculations, oxidation rate calculations, and air stoichiometry calculations and, as a result, errors or inaccuracies in the flow rate measurement impact these other variables as well. Therefore, accurate flue gas flow rate measurements are critical if an accurate model of oxidation rate is desired.

Oxygen partial pressure was not measured directly, and oxidation rate could not be measured directly. Because of this, oxygen partial pressure, oxidation rate, air stoichiometry, and mass-transfer coefficients were calculated from a mass balance involving flue gas flow rate, oxidizing air flow rate, SO_2 removal, and oxidation. Because all these factors could only be calculated from the same base data, it was not possible meaningfully to correlate them against each other.

Figure 9 is a plot of solid phase oxidation versus air stoichiometry. While the rate does not appear to vary with stoichiometry in any consistent manner, it is possible to draw some conclusions concerning net oxidation. It shows when gross air stoichiometry was 2.0 or greater, net oxidation was consistently 95 percent or greater. Thus, an air stoichiometry of 2.0 to 2.1 could be considered a conservative guideline for this agitator/sparge ring configuration.

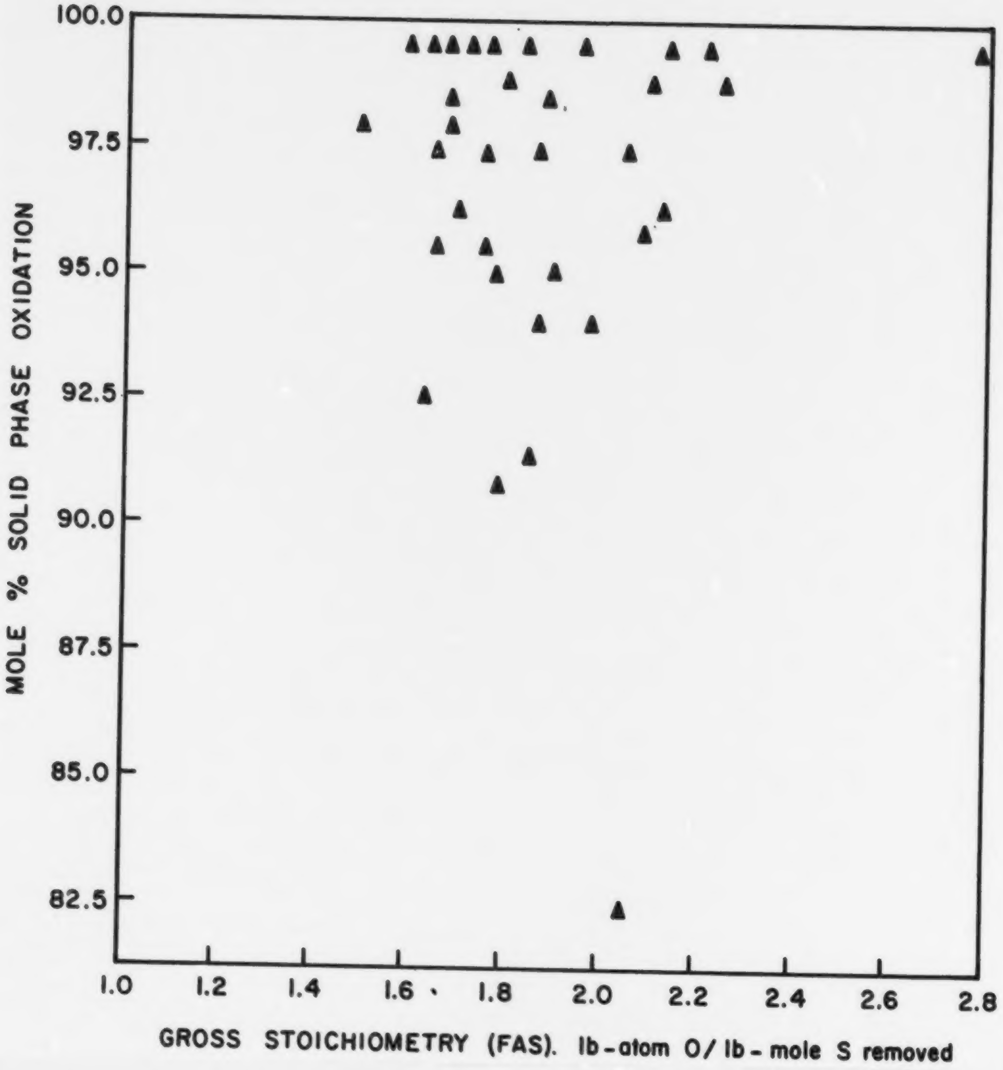


Figure 9. Solid Phase Oxidation vs Air Stoichiometry

The analysis in the preceding section suggests that the oxidation reaction is O_2 mass-transfer limited. Although the data set was not suited to formulating a generalized model, mass-transfer principles identify several factors that will affect the process. Air sparging rate and agitation are both very important mass-transfer parameters, but increases in either parameter will result in increased capital and operating costs.

RESULTS AND CONCLUSIONS

The following results and conclusions were reached at the conclusion of this demonstration program. The process areas covered include system chemistry, the oxidation reaction, and oxidation and dewatering equipment. Most of these conclusions are specific to dual-tank air sparging at Widows Creek.

- The reaction appears to be limited by the rate of oxygen mass transfer. Neither bisulfite dissolution, reaction kinetics, nor gypsum precipitation is the rate-limiting step.

- During testing with air sparged in both hold tanks (C series), SO_2 pickup in the scrubber was the primary source of bisulfite ions for the oxidation reaction. When only the venturi tank is sparged, solids dissolution is a necessary link in the reaction pathway, though not necessarily a rate limitation.

- During dual-tank oxidation tests, bisulfite ion availability was not a rate-limiting factor, and the reaction was relatively insensitive to pH. The pH should be maintained below 6.5 to keep sufficient bisulfite available for the reaction.

- SO_2 absorption in the scrubber is sufficient to maintain the pH in the proper range when oxidation is performed within the scrubber loop.

- Forced oxidation did not cause gypsum scaling. Gypsum relative saturation was only slightly greater than one during these tests, due to high availability crystal surface area. Gypsum crystals existing on packing will tend to grow in both oxidized and unoxidized systems, resulting in scaling of the packing. Forced oxidation will not eliminate maintenance or cleaning requirements for scrubber internals.

- Forced oxidation nearly eliminated solid calcium sulfite in the FGD waste at Widows Creek. For solids samples that met the 95 percent oxidation criterion, mean calcium sulfite was less than 1 percent by weight.

- The gypsum particles fell primarily in the 20 to 100 µm size range.

- The results of testing could not be modeled accurately, largely because of inaccurate estimates of the flue gas flow rate.

- An air stoichiometry of 2.0 lb-atoms O/mole SO_2 absorbed provided consistent sulfite oxidation of $\geq$95 percent.

- Calculated thickener unit-area requirements for oxidized sludge are 1.7 square feet per ton per day of solids or less. A more conservative figure should be used in design. Settling-test results showed the compression point was 40 percent solids or greater.

- The small-scale thickener at Widows Creek was underloaded, and had an unreliable rake mechanism. Its operation does not necessarily predict the operation of a full-scale unit.

- The filter-sizing criterion for Widows Creek oxidized sludge is approximately 200 pounds of solids per hour per square foot of cloth, based on industry experience.

- The small-scale filter at Widows Creek was underloaded and its operation does not necessarily predict the operation of a full-scale unit.

- Filter cake product of 75 to 85 percent solids was attained with forced oxidation.

- Forced oxidation did not significantly affect SO_2 removal.

- pH is too insensitive to dissolved limestone concentrations to be an effective control point for limestone feed rate.

- Forced oxidation can result in increased total dissolved solids (TDS) in the scrubbing liquor as a result of water reuse.

- The effects of increased TDS on the Widows Creek system are not presently known and should be studied prior to applying forced oxidation to these units.

REFERENCES

1. Dean, John A., Editor. Lange's Handbook of Chemistry, Eleventh Edition. McGraw-Hill.

2. Levenspiel, Octave. Chemical Reaction Engineering, 2nd Ed., New York. John Wiley and Sons. 1972.

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The contents of this paper do not necessarily reflect the view and policies of the Tennessee Valley Authority, nor does mention of any trade names or commercial products constitute endorsement or recommendation for use.

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LA CYGNE STATION UNIT NO. 1
WET SCRUBBER OPERATING EXPERIENCE

by

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In the late 1960's, Kansas City Power & Light Company and Kansas Gas & Electric Company entered into a joint venture to construct an 800 MW coal fired generating unit. An east central Kansas location was selected for its ample coal reserves and adequate water supply capabilities.

The coal reserves proved to be a low grade with an average of 5¼ percent sulfur and 24 percent ash. To make this coal an acceptable boiler fuel a large scale air quality control system was required. After considerable pilot testing on a smaller generating unit burning similar coal, a venturi - absorber scrubber using limestone as the scrubbing agent was selected. Construction of the La Cygne Station Unit #1 started in April 1969 and began commercial operation in June 1973.

This paper presents a review of the operating experiences, O & M cost trends, availabilities, modifications, manpower and other supportive data relating to this limestone scrubber system.

STATION DESCRIPTION

The 820-megawatt La Cygne No. 1 Unit began commercial operation on June 1, 1973, as a joint project of Kansas Gas and Electric Company and Kansas City Power and Light Company. The companies share equally in ownership and output, and the unit is operated by KCP&L. The 630-megawatt No. 2 Unit, in service since being declared commercial May 15, 1977, operates under an identical arrangement.

The plant site is located about 55 miles south of downtown Kansas City, one-half mile west of the Missouri State line, and was selected based on locally available coal, water, and limestone. Construction of No. 1 Unit began in 1969 and erection of the Air Quality Control System was initiated in mid 1971.

Water for cooling purposes is furnished from a 2,600 acre reservoir constructed adjacent to the plant site. Fly ash and spent slurry from the AQC system is piped to a 200 acre settling pond located east of the reservoir.

Coal is delivered to the plant in off-the-road 120 ton trucks from surface mines operated by the Pittsburg & Midway Coal Mining Co. The nearby coal deposits are estimated to contain 70 million tons. The fuel is low grade, sub-bituminous with an as-fired heating value of 9,000 to 9,700 Btu/lb. and an ash content of 25 per cent and sulfur content of 5 to 6 per cent (Table 1)

Limestone is obtained from nearby quarries and delivered to the plant in off-the-road 50 ton trucks.

The boiler for No. 1 Unit is a cyclone-fired, supercritical, once-through, balanced-draft Babcock & Wilcox unit, with a rating of 6,200,000 pounds of steam per hour, 1,010 degrees F, 3,825 psig at the superheat outlet. The turbine-generator was supplied by Westinghouse and is rated at 874 MW gross output with five per cent overpressure and 3,500 psi throttle pressure. Three auxiliary, oil-fired boilers are used for plant start-up or for powering a 20 megawatt house turbine-generator. The net plant output is 820 megawatts, adjusted to include 24 megawatts used by the FGD system and 30 megawatts by plant auxiliaries.

FGD SYSTEM DESCRIPTION

The La Cygne wet limestone FDG system (AQC) consists of eight venturi-absorber modules, connected together by a common inlet and outlet plenum, capable of treating 2,760,000 ACFM of boiler flue gas at 285°F. The ductwork design is such that flue gas cannot bypass the system, but

Table 1

LA CYGNE STATION
COAL AND ASH ANALYSIS

COAL

Proximate	
Volatile	28.63
Fixed Carbon	37.94
Ash	24.36
Moisture	9.07
	100.00
BTU/lb.	9421
Grindability	59.59

Ultimate	
Moisture	8.60
Carbon	51.93
Hydrogen	3.43
Nigrogen	0.94
Chlorine	0.027
Sulfur	5.39
Ash	24.36
Oxygen	5.33
	100.007

ASH

Analysis	
Phosphorous Pentoxide	0.15
Silica	46.05
Ferric Oxide	19.23
Alumina	14.07
Lime	6.86
Magnesia	1.02
Sulfur Trioxide	7.85
Potassium Oxide	2.48
Sodium Oxide	0.60
Titania	1.02
Other	0.67
	100.00

Fusion Temperature	
Reducing I.D.	1957
Soft (H=W)	2045
Soft (H=W/2)	2169
Fluid	2321
Oxidizing I.D.	2156
Soft (H=W)	2338
Soft (H=W/2)	2415
Fluid	2520

each individual module can be isolated from the system for maintenance.

The on-site limestone processing facility is composed of two 110 ton/hr. wet ball mills and two 260,000 gal. storage tanks, capable of supplying up to 1,000 tons of slurry per hour from 3/4 in. x 0 in. limestone. This slurry is processed to consist of 20 per cent solids by weight.

The unit is a balanced draft system with three 7000 hp forced draft fans and six induced draft fans located between the AQC and the 700 foot stack.

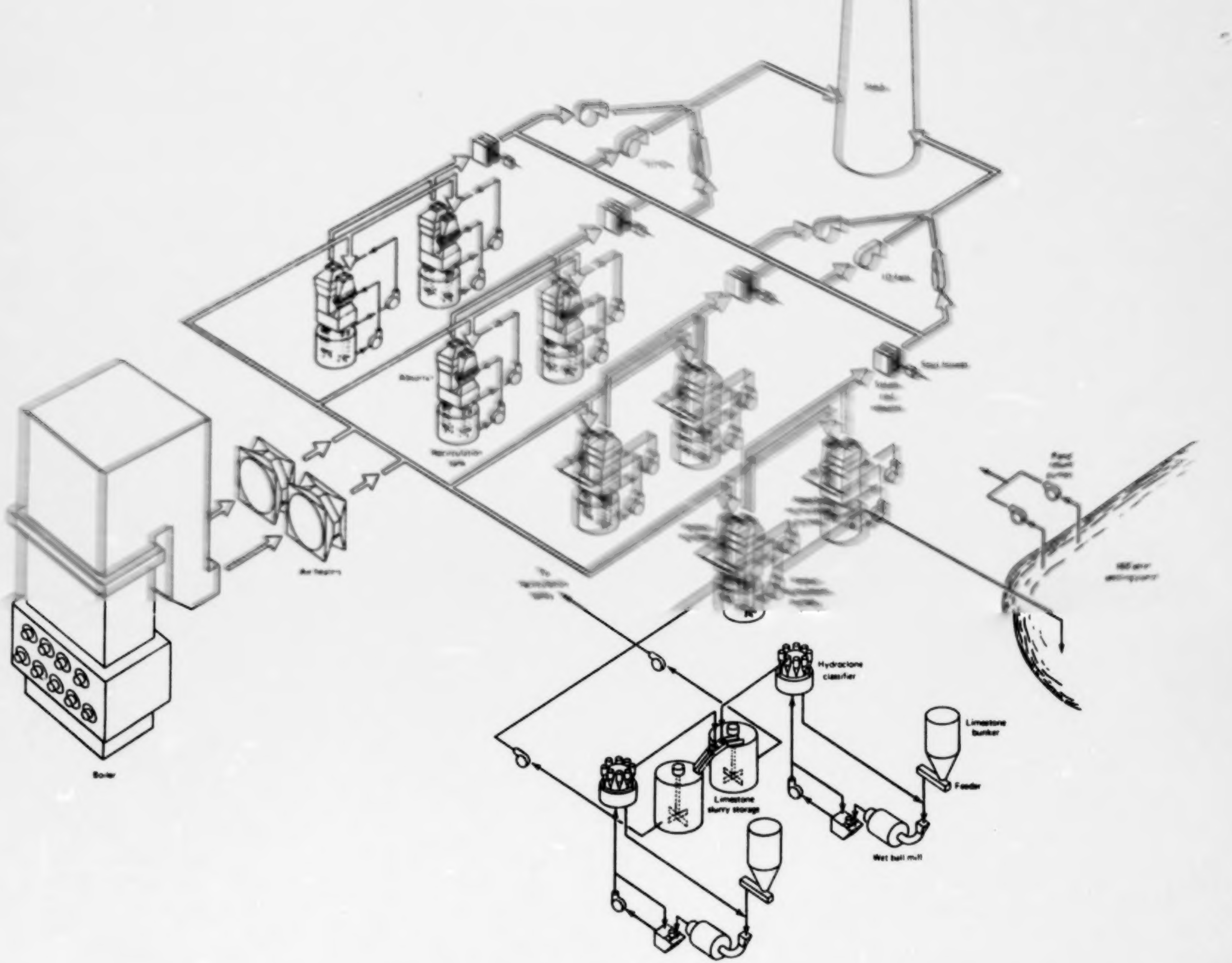
The spent slurry and fly ash is removed from the module recirculation tanks thru rubber lined pipe to a 300 acre settling pond at a rate of 3000 to 3500 tons per day. Clear make-up water is pumped from this pond for slurry make up, sump pump operation, and wash water thus allowing a closed loop operation. (Figure 1)

The hot boiler flue gas first enters the venturi section (Figure 2) of the module and is sprayed with limestone slurry in a concurrent manner from 48 spray and 32 wall wash nozzles. This results in agglomeration of up to 99 per cent of fly ash particles which is collected in the sump below. The flue gas then makes a 180 degree turn up through two layers of stainless steel sieve trays upon which slurry is sprayed from 24 spinner vane nozzles. At this point the SO_2 in the flue gas and the calcium carbonate in the slurry react to form two relatively insoluable salts, calcium sulfite and calcium sulfate, which also fall to the sump. The scrubbed flue gas then passes thru a series of demisters and is then reheated before entering the induced draft fans.

OPERATING EXPERIENCE

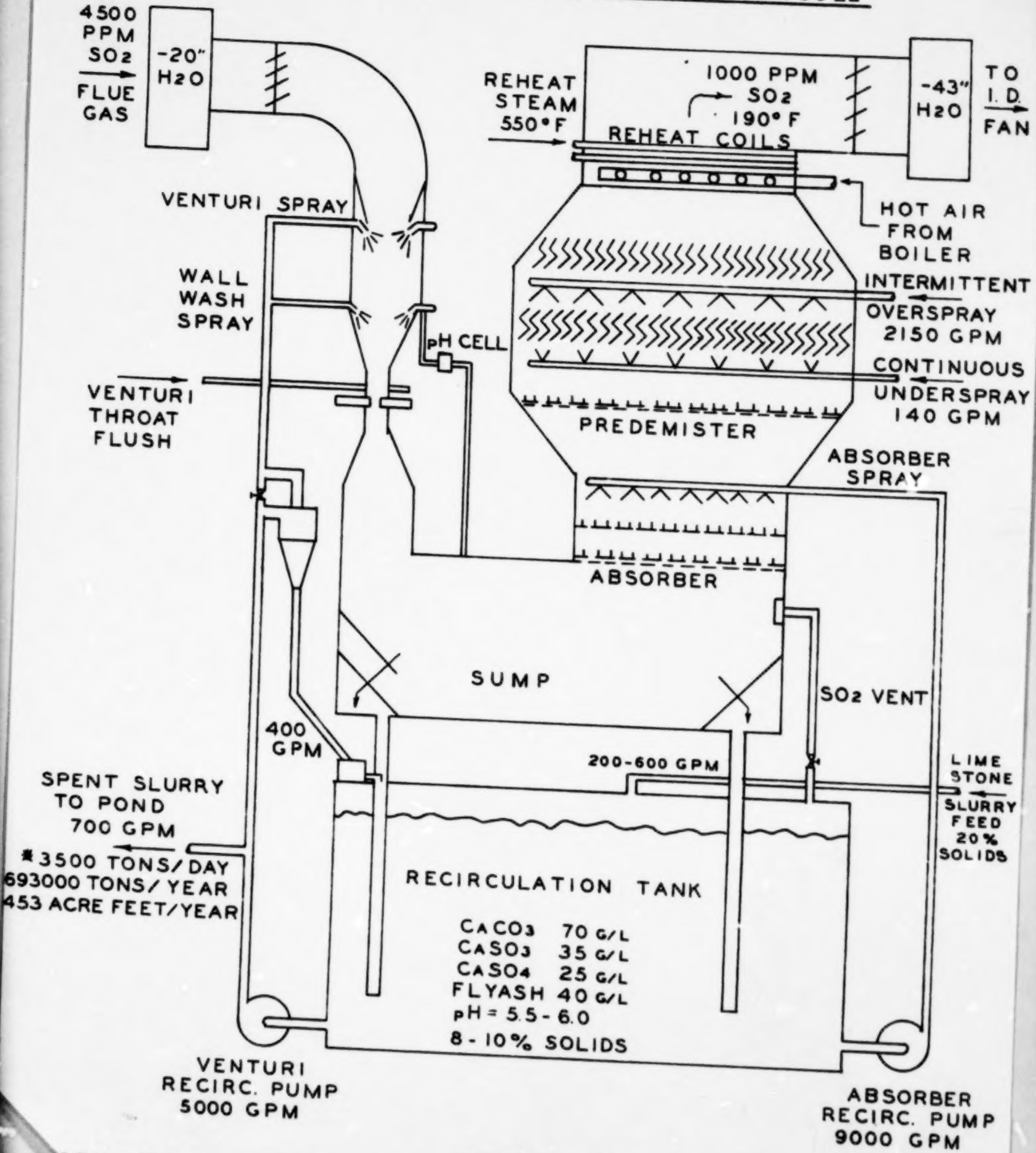
As a result of the continuing modifications and improved operating procedures, the module availabilities have steadily improved. The annual averages (Table 2) have been 31% for 1973, 76.3% for 1974, 84.3% for 1975, 92% for 1976, 92.5% for 1977, 93.5% for 1978, and 95.1% for 1979. With the addition of the eighth module in April 1977, continuous daytime load capability has exceeded 800 megawatts without appreciably affecting average module capability.

The results of a full load and stack emissions test on August 26, 1977, (Table 3) indicated module gas flow was still below cruising capability, the induced and forced draft fans were loaded well below rating and most systems were in good balance. Sulfur dioxide removal efficiency


Stack
Boiler
Air heaters
Recirculation tank
To recirculation tanks
Hydroclone classifier
Limestone bunker
Feeder
Wet ball mill
Limestone slurry storage

La Cygne limestone wet scrubbing system

Figure 1


LACYGNE FGD MODULE

4500 PPM SO2
FLUE GAS
-20" H2O
VENTURI SPRAY
WALL WASH SPRAY
VENTURI THROAT FLUSH
pH CELL
REHEAT STEAM 550°F
1000 PPM SO2 190°F
REHEAT COILS
-43" H2O
TO I.D. FAN
HOT AIR FROM BOILER
INTERMITTENT OVERSPRAY 2150 GPM
CONTINUOUS UNDERSPRAY 140 GPM
PREDEMISTER
ABSORBER SPRAY
ABSORBER
SUMP
SO2 VENT
400 GPM
200-600 GPM
LIME STONE SLURRY FEED 20% SOLIDS
SPENT SLURRY TO POND 700 GPM
*3500 TONS/DAY
693000 TONS/YEAR
453 ACRE FEET/YEAR
RECIRCULATION TANK
CaCO3 70 G/L
CaSO3 35 G/L
CaSO4 25 G/L
FLYASH 40 G/L
pH = 5.5 - 6.0
8 - 10% SOLIDS
VENTURI RECIRC. PUMP 5000 GPM
ABSORBER RECIRC. PUMP 9000 GPM

OTAL FOR ALL MODULES

Figure 2

80%

Table 2

MODULE AVAILABILITY SUMMARY - 1973

MONTH	A	B	C	D	E	F	G	AVERAGE % AVAILABILITY*	NET MWH	BOILER HOURS	GENERATION LOAD FACTOR
JANUARY											
FEBRUARY											
MARCH											
APRIL											
MAY											
JUNE	20	21	40	21	27	30	23	26	87,529	294	15.2
JULY	7	24	25	41	27	25	31	26	90,669	303	15.2
AUGUST	79	64	65	74	47	48	70	64	250,319	699	42.1
SEPTEMBER	13	0	13	13	13	0	0	7	20,073	95	3.5
OCTOBER	28	41	34	54	33	3	46	34	117,106	452	19.7
NOVEMBER	48	1	38	4	63	59	49	37	104,255	463	18.1
DECEMBER	42	20	5	31	26	11	32	24	61,013	339	10.3
								31%			17.7%

*$\frac{\text{MODULE HOURS}}{\text{HOURS IN MONTH}}$

Table 2 (Cont'd)

MODULE AVAILABILITY SUMMARY - 1974

MONTH	A	B	C	D	E	F	G	AVERAGE % AVAILABILITY*	NET MWH	BOILER HOURS	GENERATION LOAD FACTOR
JANUARY	49	32	44	87	23	37	81	50	35,862	364	6
FEBRUARY	66	68	59	76	52	100	65	69	85,256	364	16
MARCH											
APRIL	67	70	75	88	74	100	88	80	83,880	332	15
MAY	69	83	78	85	78	84	80	80	157,949	500	27
JUNE	92	84	83	90	82	83	87	86	185,473	480	32
JULY	75	80	80	81	85	79	77	80	110,122	313	19
AUGUST	90	90	73	81	81	78	99	85	231,382	571	39
SEPTEMBER	69	88	73	76	83	89	86	81	209,127	606	36
OCTOBER	71	61	59	81	79	93	89	76	230,302	662	39
NOVEMBER	90	71	60	61	84	85	84	76	130,128	386	23
DECEMBER											
								76.3%			25%

*MODULE HOURS / BOILER HOURS

Table 2 (Cont'd)

MODULE AVAILABILITY SUMMARY - 1975

MONTH	A	B	C	D	E	F	G	AVERAGE AVAILABILITY*	NET MWH	BOILER HOURS	GENERATION LOAD FACTOR
JANUARY		Turbine Generator Repair									
FEBRUARY		Turbine Generator Repair							7,886		
MARCH	82.4	96.0	89.5	76.6	93.0	91.5	96.0	89.33	244,873	694	41.1
APRIL		Generator Repair 25 Days							23,014		3.4
MAY	94.6	85.1	94.2	89.5	89.8	89.3	83.4	89.4	332,526	683	55.9
JUNE	87.8	85.4	83.9	84.9	84.1	86.1	88.6	85.8	324,952	667	56.4
JULY	78.4	89.7	89.6	83.7	85.4	87.4	85.2	85.6	297,870	590	50.0
AUGUST	74.5	88.1	87.3	78.0	92.4	85.0	83.1	84.07	294,402	630	49.5
SEPTEMBER	78.4	83.6	84.4	84.7	78.8	77.8	74.2	80.25	239,954	610	41.7
OCTOBER	66.2	77.3	46.3	73.6	71.9	73.1	64.7	67.57	74,660	231	12.5
		Generator Repair 15 Days									
NOVEMBER	92.9	90.8	80.2	93.2	96.1	89.4	93.9	90.83	165,058	346	28.7
		Generator Repair 17 Days									
DECEMBER	90.7	87.4	80.9	85.2	86.9	88.6	83.7	86.19	278,597	597	46.8
								84.3			38.6

*WORKING HOURS + RESERVE
HOURS IN MONTH

Table 2 (Cont'd)
MODULE AVAILABILITY SUMMARY - 1976

MONTH	A	B	C	D	E	F	G	AVERAGE AVAILABILITY*	NET MWH	BOILER HOURS	GENERATION LOAD FACTOR
JANUARY	85.8	84.6	90.7	71.8	83.9	82.3	84.3	83.3	301,641	620.5	50.6
FEBRUARY	93.9	90.3	85.8	91.2	91.7	93.1	94.6	91.5	308,361	594.5	55.4
MARCH	92.3	89.7	88.4	93.0	94.2	91.3	91.4	91.5	337,468	643.0	56.7
APRIL	92.3	90.5	88.7	97.1	95.8	98.0	94.8	93.9	76,810	143.0	13.3
		Scheduled Outage 24 Days									
MAY	96.5	92.1	93.5	95.7	89.4	95.3	96.2	94.1	223,048	436.3	37.5
		Scheduled Outage 9 Days									
JUNE	93.3	94.1	94.0	95.0	92.3	93.5	90.6	93.3	320,701	656.0	55.7
JULY	95.6	95.0	91.9	92.9	93.0	93.7	94.0	93.7	359,028	688.3	60.3
AUGUST	94.1	93.1	91.8	93.4	91.8	90.4	87.6	91.7	275,014	521.0	46.2
		Turbine Repair, Stack Relining 8 Days									
SEPTEMBER		Turbine Repair, Stack Relining 30 Days									
OCTOBER	97.4	96.7	97.5	89.0	96.1	96.1	96.1	95.6	88,925	255.3	14.9
		Turbine Repair, Stack Relining 18 Days									
NOVEMBER	94.7	93.3	93.7	95.3	94.2	91.3	93.6	94.0	342,236	626.8	59.4
DECEMBER	86.8	88.5	81.0	93.5	93.6	94.7	91.4	89.9	358,338	706.3	60.2
								92.0			46.4

$$*\frac{\text{WORKING HOURS + RESERVE}}{\text{HOURS IN MONTH}}$$

Table 2 (Cont'd)

MODULE AVAILABILITY SUMMARY - 1977

MONTH	A	B	C	D	E	F	G	H	AVAILABILITY*	NET MWH	BOILER HOURS	LOAD FACTOR
JANUARY	94.2	90.0	95.0	95.1	94.5	91.6	89.8	---	92.9	255,822	539	43.0
FEBRUARY	93.4	93.0	92.6	93.8	93.3	93.9	88.0	---	92.5	310,748	590	57.8
MARCH	94.0	92.2	85.9	94.3	91.4	94.0	90.1	---	91.7	295,420	558	49.6
APRIL	96.1	93.7	97.0	94.2	95.2	96.1	94.5	---	95.2	178,226	384	30.9
MAY		Generator Repair And										
JUNE		Stack Relining - 63 Days										
JULY	95.0	92.8	94.4	94.8	94.6	94.9	95.4	95.4	94.6	213,334	485	35.8
AUGUST	88.9	55.2	93.2	93.1	89.7	92.8	92.9	93.3	87.4	253,605	501	42.6
SEPTEMBER	93.2	93.7	89.1	90.0	92.8	95.0	91.7	93.0	92.3	287,701	524	49.9
OCTOBER	90.7	95.6	89.3	94.2	93.4	93.5	88.5	93.0	92.3	173,979	457	29.2
NOVEMBER	93.1	96.3	93.4	94.2	92.2	92.5	95.5	95.1	94.0	118,439	234	20.6
		Turbine Repair Nov. 15 - Dec. 25										
DECEMBER												
									92.5%			39.9

*$\frac{\text{WORKING HOURS \& RESERVE HOURS}}{\text{HOURS IN MONTH}}$

Table 2 (Cont'd)
MODULE AVAILABILITY SUMMARY - 1978

MONTH	A	B	C	D	E	F	G	H	AVAILABILITY*	NET MWH	BOILER HOURS	GENERATION LOAD FACTOR
JANUARY	90.2	94.8	94.6	95.1	93.4	93.5	94.4	94.0	93.8	332,033	582	54.2
FEBRUARY	92.4	93.4	95.1	94.3	90.6	96.9	95.5	93.4	94.0	334,897	594	60.5
MARCH	95.3	95.2	90.4	95.4	94.4	94.7	88.6	93.3	93.4	264,961	593	43.2
APRIL	91.4	92.1	92.8	90.8	90.2	91.8	90.6	90.5	91.3	330,571	620	55.7
MAY	88.9	91.5	91.6	93.1	91.5	90.6	93.1	85.6	90.7	291,651	582	47.6
JUNE		OUTAGE 6-8-78 thru 7-17-78									14	0
JULY	87.9	97.2	91.9	93.9	88.4	92.8	93.1	95.3	92.6	160,847	340	26.2
AUGUST	92.1	92.5	95.0	95.7	92.7	94.3	94.7	95.3	94.0	307,378	579	50.1
SEPTEMBER	96.1	96.0	96.3	95.8	95.9	95.7	95.3	96.6	96.0	390,826	720	65.9
OCTOBER	95.9	95.5	98.3	97.0	97.0	97.6	96.7	96.3	96.8	138,126	255	22.5
NOVEMBER	91.7	94.9	94.3	93.3	93.6	93.0	94.3	96.1	93.9	386,402	720	65.1
DECEMBER	93.9	92.9	94.0	95.0	94.7	90.5	94.4	94.7	93.8	91,744	239	15
									93.5			42.2

$$^*\frac{\text{WORKING HOURS \& RESERVE HOURS}}{\text{HOURS IN MONTH}}$$

Table 2 (Cont'd)

MODULE AVAILABILITY SUMMARY - 1979

MONTH	A	B	C	D	E	F	G	H	SYSTEM AVAILABILITY*	NET MWH	BOILER HOURS	GENERATION LOAD FACTOR
JANUARY	95.6	96.5	97.2	96.3	90.7	97.2	97.2	95.4	95.8	46,538	205	7.82
FEBRUARY	95	94.6	92.6	93.5	95.1	94.3	94.1	93.8	94.1	141,322	342	26.29
MARCH	96.1	96.0	93.2	95.6	96.5	94.8	95.7	93.4	95.2	147,645	314	24.81
APRIL	95.5	95.7	94.4	91.4	95.5	96.2	95.9	95.7	95.0	342,240	638	59.43
MAY	96.5	96.3	96.7	95.3	95.4	95.7	96.3	95.5	96.0	222,924	452	37.45
JUNE		Outage May 26 - August 16										
JULY												
AUGUST	86.8	95.9	96.3	96.3	95.9	96.2	88.5	96.9	94.1	83,169	230	13.97
SEPTEMBER	96	96.1	95.6	94.3	96.7	96.1	96.0	96.9	96	321,108	618	55.75
OCTOBER	95.3	95.8	94.7	92.7	94.4	94.9	94.7	94.5	94.6	207,639	455	34.89
NOVEMBER		Outage October 19 - Dec. 31										
DECEMBER												
									95.1			32.55

*WORKING HOURS & RESERVE HOURS
PERIOD HOURS

Table 2 (Cont'd)

MODULE AVAILABILITY SUMMARY - 1980

MONTH	A	B	C	D	E	F	G	H	SYSTEM AVAILABILITY*	NET MWH	BOILER HOURS	GENERATION LOAD FACTOR
JANUARY		Outage January 1 - February 20										
FEBRUARY	98.2	98.2	97.4	99.1	98.2	99.1	99.1	99.6	98.6	52,768	157	9.48
MARCH	94.6	96.2	96.1	96.1	95.8	94.7	93.3	95.7	95.3	1,187	32	2.00
APRIL	96.3	95.1	95	96.7	95.3	92.5	97.0	97.0	95.6	206,936	472	35.93
MAY	96.4	94.6	95.7	95.9	96.1	96.5	96.1	96.8	96.0	324,478	689	55.19
JUNE	98.2	98.0	97.4	98.1	98.1	98.3	98.7	99.3	98.3	195,974	370	34.02
JULY												
AUGUST												
SEPTEMBER												
OCTOBER												
NOVEMBER												
DECEMBER												

*$\frac{\text{WORKING HOURS \& RESERVE HOURS}}{\text{PERIOD HOURS}}$

Table 3

LA CYGNE STATION UNIT NO. 1

FOUR HOUR FULL LOAD & STACK EMISSION TEST

DATE: August 26, 1977
TIME: 11:00 A.M. - 12:00 Midnight
LOAD RANGE: 800 + MW
AMBIENT TEMP: 94° F

NOX EMISSION: 0.81 # mm BTU
AVERAGE SO_2 REMOVAL: 77%
PARTICULATE EMISSION: .213 # mm BTU

MODULES	A	B	C	D	E	F	G	H
GAS FLOW INDICATED	400	350	380	400	352	380	370	366
THROAT POSITION	OPEN	OPEN	OPEN	OPEN	OPEN	OPEN	OPEN	OPEN
REHEAT TEMPERATURE	170	190	150	190	185	180	160	170
VENTURI P	5	5.5	5	5	5	5	5	5
REHEATER P	2.5	5.5	4.5	4.5	5	2.55	4.5	5.5
ABSORBER DEM. P	6.5	5.5	10	7.5	7.0	6.5	8.0	7.0
REHEAT OUTLET DAMPER POS.	50	100	96	38	100	52	100	100
ID FAN AMPS	380	420	380	400	470	470	(540 MAX)	
ID FAN INLET DAMPER POS.	42	42	32	36	36	40	(% OPEN)	
FD FAN AMPS	490	470	430				(540 MAX)	
LAB pH	5.45	5.7	5.55	5.7	5.58	5.77	5.72	5.29
SULFITE g/l	60.4	72.4	101.0	74.1	70.0	43.9	43.9	63.6
CARBONATE g/l	50.3	75.6	53.1	54.4	59.4	83.8	68.1	42.5
SO_2 EFFICIENCY %	80.0	82.1	74.9	64.3	76.4	72.1	73.1	74.8
INLET (PPM)	4600	4600	4600	4600	4600	4600	4600	4600
OUTLET (PPM)	920	825	1150	2285	1085	1285	1235	1160

Table 3 (Cont'd)

CONDENSER VAC (IN. HG)	2.5	PRIMARY SUPER GAS PRESS. (IN.H_2O)	-8
WINDBOX FURNACE DIFF. PRESS (IN.H_2O)	32	HORZ REHEAT GAS PRESS. (IN.H_2O)	-9.5
SCRUBBER OUTLET PRESS (IN.H_2O)	-39"	ECON OUTLET GAS PRESS. (IN.H_2O)	-11.5
FURNACE PRESS (IN.H_2O)	-2	FEEDWATER PRESSURE (PSI)	4200
F.D. FAN DISCHARGE (IN.H_2O)	41	THROTTLE PRESSURE (PSI)	3400
PEND. REHEAT GAS PRESSURE (IN.H_2O)	-5	THROTTLE TEMP. (^oF)	1000^o
AIR FLOW (%)	85	HOT REHEAT TEMP. (^oF)	1300
BOILER EXCESS O_2 (%)	2.2	FUEL FLOW %	68
BAROMETRIC PRESSURE (IN.Hg)	29.01	FUEL HEATING VALUE (MTB)	9800
STACK GAS TEMP (^oF)	209	FLUE GAS VOLUME (MCFM)	2998
FLUE GAS MOISTURE (%)	13.66	STACK CO_2 %	13.4
STACK GAS VELOCITY Ft/Sec	103.15	STACK O_2 %	5.4

averaged 77% with individual modules averaging from 65 to 80%. Although particulate emissions from the plant have met EPA and Kansas State requirements, research and development work continues in an endeavor to reduce further the particulate emissions from Unit #1.

Limestone utilization has greatly improved with improved pH control. In the past, it has been almost insurmountable to maintain inline glass cells without caking the limestone during shutdown or abrading the cells during operation with the high concentration of fly ash. By centralizing the pH monitoring equipment and backflushing the pH cells with water for 5 minutes every eight hours "straight line" pH is resulting in approximately 30% less limestone, better control of scaling and has eliminated one more variable which hinders analysis in other areas.

Demister pluggage or scaling is no longer a problem at La Cygne. By eliminating the intermittent wash and moving the continous wash (140 GPM) from below to above the first demister with increased number of nozzles (230 GPM), the chevrons operate "squeaking clean". Further experimentation may allow a reduction in these nozzles and perhaps sequential washing to reduce excess water.

Hard scale on the reheater tubes has been eliminated by the addition of a second layer of demisters in each of the modules. Scaling of the reheaters continues to be a problem, however, it is soft and can be removed using fire hoses. The previous hard scale required high pressure water to remove the deposits.

MAINTENANCE

Cleaning schedules continue to call for taking one module out of service each night on a rotational schedule and keeping all modules available for the daytime peak loads. This allows a complete checkout of module internals to clean steam reheater pluggage, check nozzles for debris or loose rubber pluggage, to clean sump accumulation and to inspect for any other maintenace that could reduce reliability during the week. Module inspection and cleaning is now reduced to six hours or less with reheater pluggage the greatest problem. Scaling is not one of our chief problems and we ordinarily ignore soft scale that forms on walls, on beams, or on the outside of nozzles.

Carryover to the induced draft fan blades continues to require regular washings. Each fan now requires cleaning once very four to seven days. A "spinning" process with low pressure hoses has been very effective for cleaning the spare fan while out of service. The washings are usually done on a preventative basis, but must be taken out of service if bearing vibrations exceed 12 mils.

Rubber pipe linings and rubber-lined pumps have been an increasing maintenance problem. After several years operation, some materials that haven't been modified are wearing out. Rubber linings that tear out cause damage in other piping or pumps, plug nozzles and allow the steel pipes to wear through. This problem would not have been classified as serious, but this very abrasive slurry in practically continuous operation can be detrimental in trying to attain higher module availability, so a preventative maintenance program to change the piping in critical areas has been initiated.

Corrosion of carbon steel in the ductwork, dampers, induced draft fan rotors and housings, breeching and stack liner is and will continue to be our greatest concern. A replacement program has been underway since the fall of 1979. This program began at the outlet of the demister section where the walls from this point to the reheater section were replaced with 1/4" 316L stainless steel. In the reheater section replacement continued with the duct from the reheat bundles to the module outlet dampers being replaced with a coated carbon steel. New module outlet dampers have also been installed. Future plans include replacement of all the ducting from the module outlet dampers to the induced draft fans, the induced draft fans inlet and outlet dampers, induced draft fan housings, and the ductwork from the induced draft fans to the stack.

MANPOWER REQUIREMENTS

The scrubber operating and maintenance force has been increased to 54 people by adding one electrician for a total of two and a maintenance foreman to supervise both electricians and technicians. The remaining personnel will remain the same (Table 4).

Also worth noting are the increased demands on present maintenance personnel to accumulate, record and evaluate operating data on water saturation trends, limestone utilization, draft fan wear rates, reheater bundle failures, lined pump failures, rubber lined pipe replacements, nozzle replacements, spare parts, etc. for preventative maintenance programs. The operators are also busy updating and extending operating instructions, special instructions and reviewing safety and training procedures.

Table 4
LA CYGNE AIR QUALITY CONTROL
MANPOWER REQUIREMENTS

OPERATORS PER SHIFT	
3 Attendants	13
3 Clean-up	14
1 Shift Foreman	5
1 Process Attendant (Chemist)	1
	33

MAINTENANCE	
Mechanics	8
Apprentice Mechanics	2
Welder	1
Electrician	2
Technician	2
Plant Helpers	2
Foreman	2
	19

ADMINISTRATIVE	
Superintendent	1
Engineer	1
	2
TOTAL	54

COSTS

The total cost of the FGD system to date has increased to $55.1 million or 25% of the $216.3 million total Unit #1 cost.

The production costs for the La Cygne FGD system (Table 5) in 1977 was 1.7 mills/KWH and for 1979 it was 4.9 mills/KWH. This drastic rise is due to the increase in maintenance materials to repair the "cold end" corrosion mentioned earlier. Discounting escalation, future production costs associated with the operating labor, operating materials, and maintenance labor should remain the same or trend downward while the maintenance materials will increase slightly.

CURRENT PROGRAMS AND PROJECTS

The major project concerning the FGD system, at present, is the systematic replacement of the cold end duct work mentioned previously. To help combat this corrosion problem, studies are continuing for increasing the reheat steam supply.

Analytical programs to investigate the mechanics involved with the scale formation at the various levels of the scrubber modules and collection of sub-micron flyash have been implemented. Results of these programs are currently under scrutinization.

Other areas of hopeful improvement are in the demister section with the addition of a third layer, and on line incline reheat tube cleaning.

Table 5 - LA CYGNE UNIT #1
FGD SYSTEM OPERATING EXPENSE

	1973 DOLLARS-MILLS/KWH	1974 DOLLARS-MILLS/KWH	1975 DOLLARS-MILLS/KWH	1976 DOLLARS-MILLS/KWH
OPERATING LABOR	$ 162,934 - 0.223	$ 284,541 - 0.223	$ 601,029 - 0.265	$ 683,939 - 0.229
OPERATING MATERIALS	3,480 - 0.005	67,032 - 0.053	195,926 - 0.086	415,226 - 0.139
MAINTENANCE LABOR	189,400 - 0.259	401,414 - 0.315	416,206 - 0.184	358,941 - 0.129
MAINTENANCE MATERIALS	441,737 - 0.604	335,486 - 0.263	386,397 - 0.171	93,292 - 0.031
LIMESTONE	264,514 - 0.362	780,297 - 0.613	1,256,048 - 0.554	1,717,949 - 0.574
TOTAL	1,062,065 - 1.453	1,868,770 - 1.467	2,855,606 - 1.260	3,269,347 - 1.102

	1977 DOLLARS-MILLS/KWH	1978 DOLLARS-MILLS/KWH	1979 DOLLARS-MILLS/KWH	1980(Jan. - June) DOLLARS-MILLS/KWH
OPERATING LABOR	$ 679,628 - 0.313	$ 755,500 - 0.250	$ 733,016 - 0.485	$ 331,654 - 0.423
OPERATING MATERIALS	253,662 - 0.117	453,140 - 0.150	537,172 - 0.355	69,632 - 0.089
MAINENANCE LABOR	476,724 - 0.219	414,355 - 0.137	561,624 - 0.371	295,094 - 0.376
MAINTENANCE MATERIALS	1,083,167 - 0.493*	757,951 - 0.251	4,398,066 - 2.90	2,399,063 - 3.056
LIMESTONE	1,202,005 - 0.553	1,452,792 - 0.482	1,183,169 - 0.782	289,265 - 0.368
TOTAL	3,695,186 - 1.695	3,833,738 - 1.270	7,413,047 - 4.89	3,384,708 - 4.312

*600,000 Pond Dredging

Table 6
LA CYGNE SCRUBBER WATER ANALYSIS

CATIONS	COOLING LAKE	SETTLING POND
Calcium (Ca)	126.4	808.0
Magnesium (mg)	16.3	106.0
Sodium (Na)	31.0	52.5
Potassium (K)	5.1	41.6
ANIONS		
Bicarbonate Alk (AS HCO_3)	112.2	79.3
Chloride (CI)	44.9	314.0
Sulfate (SO_4)	295.2	1995.1
Sulfite (SO_3)	* ND	* ND
Silica ($Si0_2$)	1.12	52.0
OTHERS		
pH (pH UNITS)	7.7	7.5
Conductivity in Michromhos	820.0	3500.0
Solids, Suspended	5.0	5.0
Dissolved	610.0	3450.0

*ND - Not Detected

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EPA Symposium on Flue Gas Desulfurization (1974, 1976, 1977)

and

Mr. Terry Eaton
EPA Symposium on Flue Gas Desulfurization (1979)

ONE BUTTON OPERATION
START-UP OF THE ALABAMA ELECTRIC COOPERATIVE FGD SYSTEM

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ABSTRACT

In September of 1978, Alabama Electric Cooperative started up a limestone FGD system for its 255 MW Tombigbee Station Unit #2, Leroy, Alabama. Since the start-up, the operating experience of the system has been extremely successful.

A sophisticated control system has been provided for the FGD system which permits operation from the control room by means of a single button. Start-up of the FGD system consisted of pushing this button. The unit has been on stream since that time.

The FGD system is designed to remove 85% of the SO_2 in the flue gas generator from the combustion of 1.8% coal. Under performance test conditions the absorber gas load and inlet SO_2 content were 20% and 35% respectively greater than design. Despite the greater than design conditions a SO_2 removal efficiency of 94% was achieved. A limestone stoichiometry of 1.01 was obtained, probably the lowest ever achieved in the FGD industry.

After a year of operation, the system has exhibited a high degree of reliability. Based upon actual measured hours, the system availability has been 91.6%.

The FGD system for Unit No. 3, a duplicate of Unit No. 2, has recently been started up. Preliminary results indicate similar performance to that obtained with Unit No. 2.

It is the purpose of this paper to discuss in detail the process chemistry, system description and controls which have permitted the successful operation of this unit.

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INTRODUCTION

Alabama Electric Cooperative's Tombigbee Power Station is located on the Tombigbee River approximately 70 miles due north of Mobile, Alabama. The most recent expansion at this site was the addition of Units No. 2 and No. 3. Each unit has a rated capacity of 255MW and is designed to burn Alabama and Kentucky coals with a maximum sulfur content of 1.8%. To meet the emission standard of 1.2 lbs. SO_2 per million BTU, flue gas desulfurization was required. In September 1975 Peabody Process Systems was awarded a contract to furnish a limestone FGD system for both units.

The FGD system for Unit No. 2 was started up in September 1978. Unit No. 3 was put in service July 1979. In commissioning both units, start-up was achieved by the pushing of a single button located in the control room.

The pushbutton start-up was simple. However, the ease and simplicity of the start-up was not an accident. It was the result of careful attention to process design, mechanical design and pre-commissioning check out of the system. Since the initial pushbutton start-up, superior operating results have likewise confirmed the importance of giving proper attention to these design details. In the sections that follow, the details which contributed to the success of the Alabama Electric Cooperative System will be discussed as well as the performance history for both units.

SYSTEM DESCRIPTION

The system design criteria are shown in Table No. 1.

TABLE NO. 1

DESIGN BASIS PER UNIT	
Unit Generating Capacity	255MW
Coal Sulfur Content	1.8% S.
SO_2 Emission Standard	1.2 lbs. SO_2/mm BTU
Flue Gas Volume	953,000 ACFM
Percent of Flue Gas Scrubbed	70%
No. of Absorbers	2
No. of Recycle Pumps/Absorbers	3
Absorber SO_2 Removal Efficiency	85
Alkali	Limestone
Waste Solids Disposal Method	Ponding
Reheat Method	By-Pass Gas

The flue gas entering the FGD System has been cleaned of particulate by means of a hot side precipitator. Two I.D. fans, providing the draft for both the boiler, precipitator and FGD system, are located ahead of the absorbers. It was the Owner's preference that two absorbers be used. Each absorber has a 22' diameter and is designed for 85% SO_2 removal. Seventy percent of the gas is scrubbed and 30% is by-passed for use as reheat. Each absorber consists of six spray banks through which a slurry containing calcium sulfite, calcium sulfate and unreacted limestone is sprayed countercurrent to the gas flow. The gas, as a result of being contacted with the slurry, is cleaned of SO_2. After leaving the absorption zone, entrained slurry in the flue gas is removed by means of a two stage mist elimination section. The first stage is a weeping sieve tray deluged with a chemically non-reactive slurry produced by means of hydroclones. The hydroclones are used to classify the absorber recycle slurry by particle size. Unreacted limestone particles are ten (10) times larger than the reacted product. Hydroclones permit removing the unreacted limestone from the reacted products because of the particle size difference. Decarbonated slurry is used to deluge the weeping sieve tray and thus prevent a plugging chemical reaction. Final de-entrainment, particularly of gas entrained water, is accomplished in a second stage which is a Chevron type mist eliminator. The clean gas then leaves the absorber where it is mixed with by-pass gas to provide reheat. At the ductwork juncture where the by-pass gas and the scrubbed gas meet, a mixing baffle is used to ensure a uniform gas temperature prior to entering the stack.

A single limestone preparation system is common to both units. Limestone rock is crushed on site to approximately a 3/4" size and stored in a silo. A weigh feeder conveys the limestone to a ball mill where it is ground to proper size and stored in a tank. The limestone is fed to the individual units as a 35% slurry via a recirculation loop. Limestone slurry is fed to the individual recycle tanks as required.

Per boiler, both absorbers are supported at grade and share a common recycle tank. The recycle slurry is recirculated from the recycle tank to each absorber. Each absorber has three recycle slurry pumps - one pump is dedicated to two absorber spray headers.

Waste slurry overflows from the recycle tank to a waste slurry sump which also collects all drainage and water used for system flushing. The waste slurry is then transported from the sump to a pond in which the solids are allowed to settle. The water reclaimed from the slurry is recycled back to the FGD system for reuse. The system operates on a totally closed loop water balance basis. Fresh water is added to the system to make up for losses resulting from evaporation and water bound with the waste solids.

FACTORS EFFECTING OPERATING AND MAINTENANCE COSTS

There are many factors which contribute to the success of an operating FGD system. The whole is truly the sum of its parts, and design detail, no matter how small, if ignored can adversely effect system performance. The following areas, which will be discussed, highlight some of the many features applied to Alabama Electric Cooperative which have contributed to the FGD System's successful operating experience.

The following factors can and do effect the operability of an FGD System:

1. Absorber design
2. Mist eliminator chemistry control
3. Simplicity of design
4. Materials of construction
5. Slurry piping design
6. Adaptability to actual operating conditions

Limestone FGD systems are frequently approached with the idea that reacting limestone with SO_2 is simple high school chemistry. This is very far from the truth; the chemistry is very complex. One of the unique features of this chemistry is that the reaction products produced can result in significant scaling and plugging in an absorber. Consequently, the type of absorber used is very significant. Industry experience has shown that in many systems an absorber with complicated internals, for example, tray type absorbers, packed towers, etc., offer great potential for solids to deposit thus hindering the operability of the absorber. For the Alabama FGD system, a spray tower absorber was used as a basis for the absorber design. A significant feature of the spray tower is that its internals are minimal and thus minimizes the opportunity for solids to deposit. Selection of the spray tower design significantly increases the operability of a FGD system. However, the mist elimination section still provides the complicated surfaces which can create plugging problems. In any absorber design, de-entrainment is a factor which must be considered to avoid particulate emission through the stack. A standard design mist eliminator can provide the opportunity for solids to deposit on the surfaces. This is attributed to the fact that entrained slurry from the absorber zone contains unreacted limestone. This unreacted limestone can then react with the remaining SO_2 in the flue gas and cause solids to deposit on the complex surfaces of the mist eliminator. Attempts are frequently made to avoid this plugging problem by washing with water. However, a system designed for a closed loop water balance usually does not have the necessary quantity of water available under all load conditions and all sulfur coals to adequately preclude a plugging situation in the mist eliminator.

An alternate is to control the chemistry by preventing entrained limestone from reaching the mist eliminator. This is achieved in the Alabama Electric Cooperative design by using hydroclones to remove unreacted limestone from the process slurry and then using the limestone free slurry to provide a liquid barrier below the mist elimination zone. This liquid barrier prevents entrained slurry, with the unreacted limestone, from reaching the critical mist elimination area. This technique insures that the mist elimination area operates with a non-plugging chemistry regardless of the load or sulfur content of the coal being burned. It is thus another step in improving the reliability of the system.

Hydroclones are also used to screen all of the waste solids prior to discharge from the system. This removes all of the unreacted limestone from the waste slurry such that the FGD system efficiency of limestone utilization is almost 100%

Simplicity of design is another important factor which adds to minimal maintenance. Generally, the fewer the number of components of a system the less the probability of having problems. This philosophy has been utilized in the control concept for the Alabama Electric Cooperative FGD System. Control valves in slurry service, which create both abrasion and plugging problems, have been eliminated completely. The only exception is a small limestone slurry feed control valve. The elimination of valves is made possible by employing gravity overflow where possible. An example of this is the main recycle tank and wash tank. The quantity of slurry to the spray headers within the absorber is controlled by turning off pumps rather than modulating the slurry flow. This eliminates both a control valve and the plugging of the slurry pipe line which would occur under low flow conditions.

In slurry services which require operating over a wide range of flows, various approaches are taken to prevent plugging of the slurry pipe lines.

Simplicity of slurry piping design is achieved by having the absorber recycle slurry pumps feed a dedicated spray header system. Thus, two levels of spray banks are dedicated to a single recycle pump. Pipe manifolds and isolation valves in the discharge pipe of the recycle pump are eliminated. Regulation of slurry flow to the absorber is effected on a stepwise basis by turning individual pumps on or off as required to meet emission standards based upon the actual sulfur content of the coal being burned. This concept also eliminates plugging problems due to the creation of dead pockets in slurry pipe systems and abrasion problems of valves in the discharge piping.

In the limestone feed system, limestone is circulated via a distribution loop such that regardless of how much slurry is required by the FGD system, (0 to 100% of design) the limestone transfer system will always have velocities sufficient to prevent settling of solids and the resulting plugging which would ensue. Likewise, in the waste solids transport system, a long distance between the absorber and the pond is very common. This line must also be capable of transporting varying quantities of waste solids resulting from fluctuating gas load and coal sulfur content. A plugging condition resulting from insufficient slurry velocities in the waste solids transport system will exist when less than design quantities of waste solids are produced during normal operation. An alternate is to design the system to operate on a constant velocity basis at all times and thus eliminate the plugging problem.

As operating conditions vary, the quantity of waste slurry overflowing from the recycle tank to the waste slurry sump will likewise vary. Reclaimed water from the pond is added to the waste slurry sump. The quantity of water added reflects the difference between the quantity of waste slurry produced and design capacity of the system. This insures that the transfer system has a slurry velocity sufficient to prevent settling of solids and prevent plugging under all operating conditions.

The FGD industry has evaluated many materials of construction with varying degrees of success. For the Alabama Electric Cooperative System, linings have been used extensively.

Consider, for example, the materials of construction selected for the absorber. All wetted parts of the system are subject to corrosion. In addition, the spray absorbtion zone must contend with abrasion. To effectively remove the SO_2 from the gas, all of the gas must be contacted with the slurry. To insure proper gas/slurry contacting and prevent short circuiting of the gas, the spray pattern must be designed such that the slurry impinges on the absorber wall which creates a sand blasting situation. To withstand both the corrosion and the abrasion, the spray absorber zone has a rubber lining. The internal spray headers are carbon steel, rubber lined, rubber covered, to withstand abrasion internally and externally. All connections are flanged and are rubber covered with high alloy bolting and backup rings to insure that the bolting does not destroy the integrity of the lining. Also critical is the selection of the spray nozzle material which in this case is a cast silicon carbide. The nozzle has no internals and has a minimum opening of 1" which makes it insensitive to plugging because of trash material in the system.

In the absorber area above and below the absorption zone, abrasion is not a problem and only corrosion must be considered. In these sections, a vinyl ester flake glass lining is used.

Regardless of how good the lining material selected, the liner is no better than the manner in which it is installed. Quality Control during installation becomes critical and directly effects the maintenance requirements of the system. Apparently minor details such as how the rubber lining sections are lapped can effect the success of the lining installation. For the Alabama design, where two dissimilar lining materials are joined, a full body flange on the absorber module is used to mechanically join the dissimilar materials. Though more expensive, the mechanical joint eliminates the problems associated with chemically bonding dissimilar materials. Chemical bonding has generally proved unsuccessful and will create a maintenance problem.

In any FGD system, the design condition specifications rarely reflect the actual operating conditions of the plant. Sulfur contents in the coal vary and load conditions vary. The absorber design provided has no limitations with regard to minimum gas flows and yet has the capability of achieving minimum operating costs by turning off recycle pumps when less than design sulfur coals are burned. This permits achieving the desired SO_2 emission level at the lowest possible operating cost.

OPERATING RESULTS

Various concepts have been discussed with regard to ensuring the reliable performance of the Alabama Electric Cooperative FGD Systems. Operating results are the proof as to how successful these concepts have been. Therefore, it is important to review actual performance.

In September of 1978, Unit #2 was placed on stream. In July of 1979 Unit #3 was placed in service. The operating results which are discussed here reflect the experience which Alabama Electric Cooperative has had over more than twenty (20) months of FGD system operation.

Start-up

The design concept Alabama Electric Cooperative chose for its control system is a fairly sophisticated one. By means of programmable controllers the total start-up and shutdown sequence of the FGD system is accomplished by the pushing of a single button. As part of the start-up, the total system had been checked out mechanically and electrically such that all subsystems were proven. Having done this, the units were started up on flue gas by means of pushing that single button. The single button start-up was achieved for Unit #2 and duplicated for Unit #3 ten (10) months later.

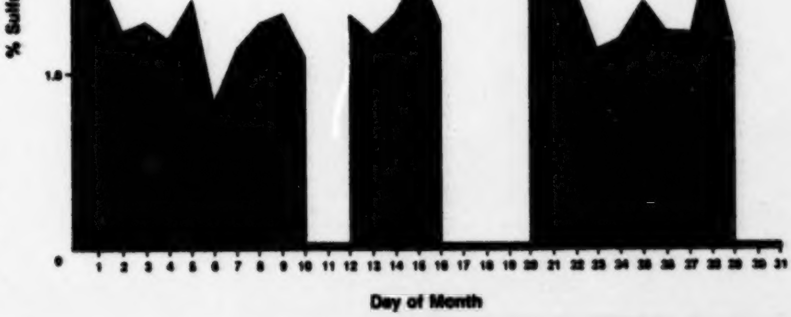
The coals which the Tombigbee Station burn are from four or five different mines located in Alabama and Kentucky. Though the maximum design was a 1.8% sulfur coal, the actual sulfur content of the coals being fired range from 0.88% to 3.6% (see Table No. 2).

TABLE NO. 2

TYPICAL COALS BURNED

PROXIMATE ANALYSIS			
MOISTURE %	5.82	7.40	4.64
ASH - %	13.98	15.10	12.62
VOLATILE MATTER - %	31.61	31.40	35.32
FIXED CARBON - %	48.60	46.11	47.42
SULFUR - %	0.88	1.06	3.62
HEATING VALUE - BTU/lb.	11,805	11,169	12,199

The sulfur content variation experienced during the month of December 1979 can be considered typical. This is shown in Figure No. 1.



SO_2 Removal and Limestone Utilization

As indicated earlier, the absorbers had been designed for 85% SO_2 removal while burning a 1.8% sulfur coal. A limestone stoichiometry of 1.10 mols of calcium carbonate/mols of SO_2 absorbed had been guaranteed. Performance tests were performed by the Owner's engineers, Burns & McDonnell. For Unit #2, a 93.5% SO_2 removal was obtained while burning a 2.7% sulfur coal at an absorber gas flow 24% greater than design. When the test was repeated for Unit #3 - a 97% removal efficiency was obtained while burning a 2% sulfur coal. In both tests, summarized in Table No. 3, a limestone utilization was very close to the theoretical quantity, which reflects 100% utilization of the limestone. This is attributable to the use of the hydroclones for removal of limestone from the waste slurry.

TABLE NO. 3

PERFORMANCE TEST RESULTS

	Design Values	Unit No. 2		Unit No. 3	
		Test Value	% of Des.	Test Value	% of Des.
% S in Coal	1.8	2.7	+50%	2.0	+11%
Gas Volume/Absorber-ACFM	270,000	335,000	+24%	270,000	+ 1%
Inlet SO_2 Conc.-ppm	1106	1614	+46%	1250	+13%
Outlet SO_2 Conc.-ppm	166	105		36	
% SO_2 Removal	85	93.5		97.1	
Limestone Stoichiometry (moles $CaCo_3$/ moles SO_2 Absorbed)	1.10	1.01		1.02	

Power

The power requirements for the system are low and are summarized in Table No. 4. The power consumption shown for the FGD system under design conditions is less than 1% of rated generating capacity. However, the capability of the system to save power when operating at less than design sulfur coals (1.1% S normal vs 1.8% S design) is evidenced by the fact only 0.6% of rated generating capacity is required.

TABLE NO. 4

FULL LOAD POWER REQUIREMENTS

	Design % S Coal	Normal % S Coal
No.of Absorbers/Unit	2	2
No. of Operating Recycle Pumps/Absorbers	3	1
Power - FGD and I.D. Fan		
KW Required/Unit	4275	3496
% of Rated Capacity	1.68	1.37
Power - FGD Only*		
KW Required/Unit	2342	1564
% of Rated Capacity	0.92	0.61

*Includes Flue Gas Pressure Drop for FGD System

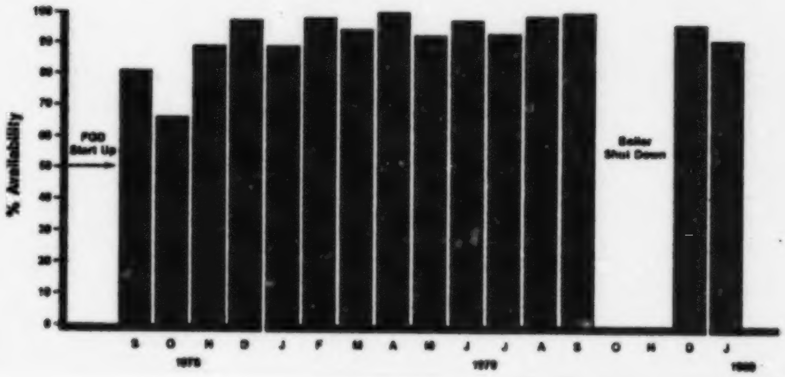
Manpower

The manpower requirements for the two operating FGD units (total 510 MW) are low - very low. Alabama Electric Cooperative employs two operators per shift, on a four shift basis. With regard to maintenance, all work is performed on a work release basis. In terms of maintenance manhours actually expended, 40 hours per week are required for instrumentation and 20-30 hours per week for mechanical work is required. These numbers are contrary to the 50 or 60 operators which are frequently cited for FGD systems.

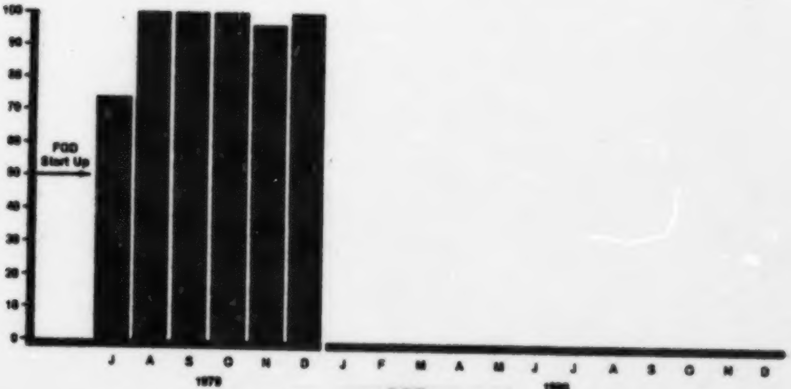
Availability

Since start-up, both units 2 and 3 have been characterized by high availability. As illustrated by the graphs shown below, as the operators learn to run the system properly, the availability, improves significantly and availabilities of 90% or greater have been consistently achieved. With the learning experience having already been gained on Unit #2, start-up of Unit #3 was virtually trouble free and this is reflected in the high availabilities achieved right from the start. Except for the first month of start-up, Unit #3 has consistently achieved monthly availabilities in excess of 97%.

Alabama Electric Cooperative
Tombigbee Station
Unit No. 2 Availability



Alabama Electric Cooperative
Tombigbee Station
Unit No. 3 Availability



PROBLEM AREAS

Like any system, problems have been experienced in the operation of the FGD system. Fortunately, the problems encountered were correctable and non-recurrent.

The ball mill system had two problem areas. A seal at the feed end of the ball mill had not been installed - this resulted in leakage of the limestone slurry. Likewise a ball retention helix at the discharge of the ball mill was not sufficiently deep to retain large particles. Peabody corrected both problems and the system now functions adequately.

No system would be complete without damper problems. In the Tombigbee Station these were also encountered. Double guillotine dampers were used in which seal air was injected between damper blades. A spare blower was provided for each seal air system. When the unit was started up two problems were encountered. The operators were undersized and would not move the damper. When the dampers were in the open position, flue gas containing SO_2 would leak into the seal air blower system, condense and create a corrosion problem. The operator problem was corrected by installing larger motors. The seal air system was corrected by installing an isolation valve between the blowers and the damper such that flue gas would not flow back into the blower system and condense. With these problems corrected, the dampers are operating satisfactorily.

Two problems were encountered in the instrumentation area. The pH sensing probe is emersed in the slurry in the absorber recycle tank. Problems were encountered with slurry leaking into the preamplifier which caused failure on several occasions. The preamplifier was changed to a different type which was enclosed in a seal housing which prevented leakage. This eliminated the problem. Gas flow measurement by means of an anubar was a total failure. Under low flow conditions, it was not possible to get a meaningful signal. Measurement of gas flow to an absorber was not critical and therefore attempts at this measurement were abandoned.

Trash material has caused the spray wash nozzles under the interface tray to plug. Placing an in-line strainer in the suction of the wash pump which feeds the spray nozzles has eliminated this problem.

The slurry transfer line from the waste sump to the pond is made of FRP pipe. Rupture of this line has occured several times because of inadequate pipe supports and also water hammer resulting from switching waste slurry pumps on and off. Pipe supports have been redesigned. The method of operating the waste slurry pumps has been modified by inclusion of a timer to provide a delay time when switching from an operating waste slurry pump to a spare pump. The object of this is to minimize the effect of water hammer.

SUMMARY OF OPERATING AND MAINTENANCE EXPERIENCE

The operating experience of Alabama Electric Cooperative has been unique and is characterized by:

1) Push button start-up
2) High availability
3) High SO_2 removal efficiency
4) High limestone utilization
5) Low manpower requirements
6) Low maintenance costs

OPERATION AND MAINTENANCE EXPERIENCE OF THE WORLD'S LARGEST SPRAY TOWER SO_2 SCRUBBERS

BY: ROBERT A. HEWITT - TEXAS UTILITIES SERVICES, INC.
and A. SALEEM - CHEMICO AIR POLLUTION CONTROL CORP.

The 750 MW Monticello boiler #3 of Texas Utilities Services, firing lignite coal, is equipped with three large spray towers, designed by Chemico Air Pollution Control Corporation. Each spray tower is sized to handle over one million cubic feet per minute of flue gas. This flue gas desulfurization system uses pulverized limestone slurry for scrubbing and includes a flue gas bypass as well as external steam flue gas reheat system. The FGD system went into operation in mid 1978 and has since logged consistently very high availability as well as high SO_2 removal efficiency. The extreme simplicity of the spray tower system has resulted in only modest increase in the power plant's operating and maintenance staff. A recent inspection of the system revealed no major problems with the tower and duct liners or the tower internals. A few isolated spots on the internal slurry pipes showed wear due to close proximity to the sprays. Failure of the rubber lining on the side mounted agitators and slurry recycle pumps has been the primary source of problems with the system. The experience with this system in general has been very satisfactory and Texas Utilities has purchased two essentially duplicate systems for the Twin Oak Power Station.

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INTRODUCTION

Texas Utilities Company is investor owned and includes three electric utilities, two fuel companies, a generating company, a service company and two non-utility companies engaged in energy related activities. The total generating capability at the end of 1979 was 17,430 megawatts. Units utilizing Texas lignite as a fuel account for 5300 megawatts of this capability and in 1979 about 50% of the total generation of the system was from the lignite fired units.

Monticello #3 is a lignite fired unit rated at 750 megawatts located at a site near Mt. Pleasant, Texas. A typical fuel analysis is shown in Table I. Units 1 and 2 were placed in service in 1974 and 1975. No flue gas desulfurization (FGD) systems were required for these units. Emission regulations applicable to the #3 unit are a maximum 2 hour average particulates emission of 0.1 lbs per 10^6 BTU, maximum opacity of 20% and a maximum 2 hour average SO_2 emission of 1.2 lbs per 10^6 BTU.

After evaluation of bids Chemico Air Pollution Control Corporation was awarded a contract to supply the electrostatic precipitators, I.D. fans and ductwork to the chimney along with major engineering and design for the SO_2 removal system with an option of provision of the total FGD system. Included in this contract was the construction and operation of a 4000 ACFM pilot plant utilizing flue gas from one of the existing units. The objectives of the pilot plant study were to determine:

a. Reactivity of available limestone.

b. Optimum stoichiometry, L/G and recycle solids.

c. SO_2 removal efficiency at full and partial load.

d. Limestone consumption.

e. Susceptability of system to plugging and scaling.

TABLE 1

TEXAS UTILITIES–MONTICELLO NO. 3
TYPICAL LIGNITE FUEL CHARACTERISTICS

Proximate Analysis–%	Typical	Range
MOISTURE	32.68	27.5–36.3
VOLATILE MATTER	30.17	25.0–34.7
FIXED CARBON	23.88	20.8–27.5
ASH	13.27	6.4–18.9
SULFUR	.72	.36–1.79
BTU AS RECEIVED	6689	6068–7302

In February, 1976 the option of having Chemico supply the necessary additional design, engineering and material for the complete FGD system was taken. The system was to be completed and ready for trial operation by January of 1978.

This FGD system is unique in that it involves the world's largest spray towers. The system has been in service since August 1978 and performance and availability have been satisfactory. The main focus of this paper is to review the operating and maintenance experiences to date.

FGD SYSTEM DESCRIPTION

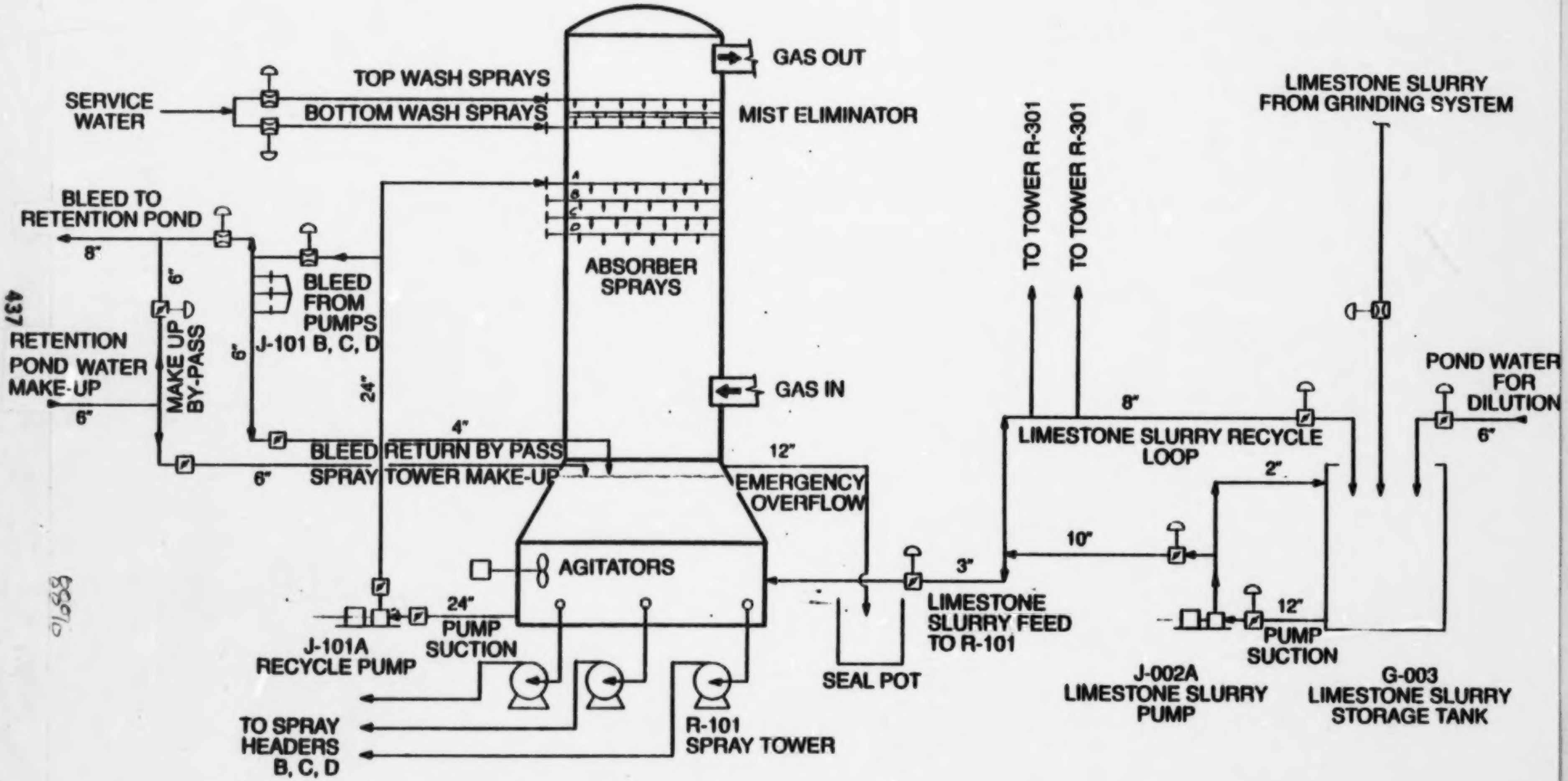
The lignite fired boiler generates about 3.4 million ACFM of flue gas at full load. The FGD system is designed to keep the SO_2 emissions to less than 1.2 lbs/MMBTU. The SO_2 removal is accomplished by scrubbing with an aqueous slurry of pulverized limestone in three large spray towers. The spent slurry containing calcium sulfur salts is disposed of in an onsite pond from which the reclaimed water is recycled to the FGD system. A simplified flow diagram of the FGD system is shown in Figure 1. The general arrangement of the major equipment is shown in Figures 2 and 3.

After passage through the electrostatic precipitators for particulate removal, three centrifugal boiler I.D. fans drive the flue gas into a common inlet manifold from which it is equally distributed into three spray towers for SO_2 removal. The scrubbed gas leaving the spray towers is again collected into a common outlet manifold for discharge into the stack.

Partial or full bypass of flue gas around the scrubbers is possible with two bypass ducts which are equipped with split louvre dampers for gas flow control. The partial bypass, up to a maximum of 50%,

FIGURE 1

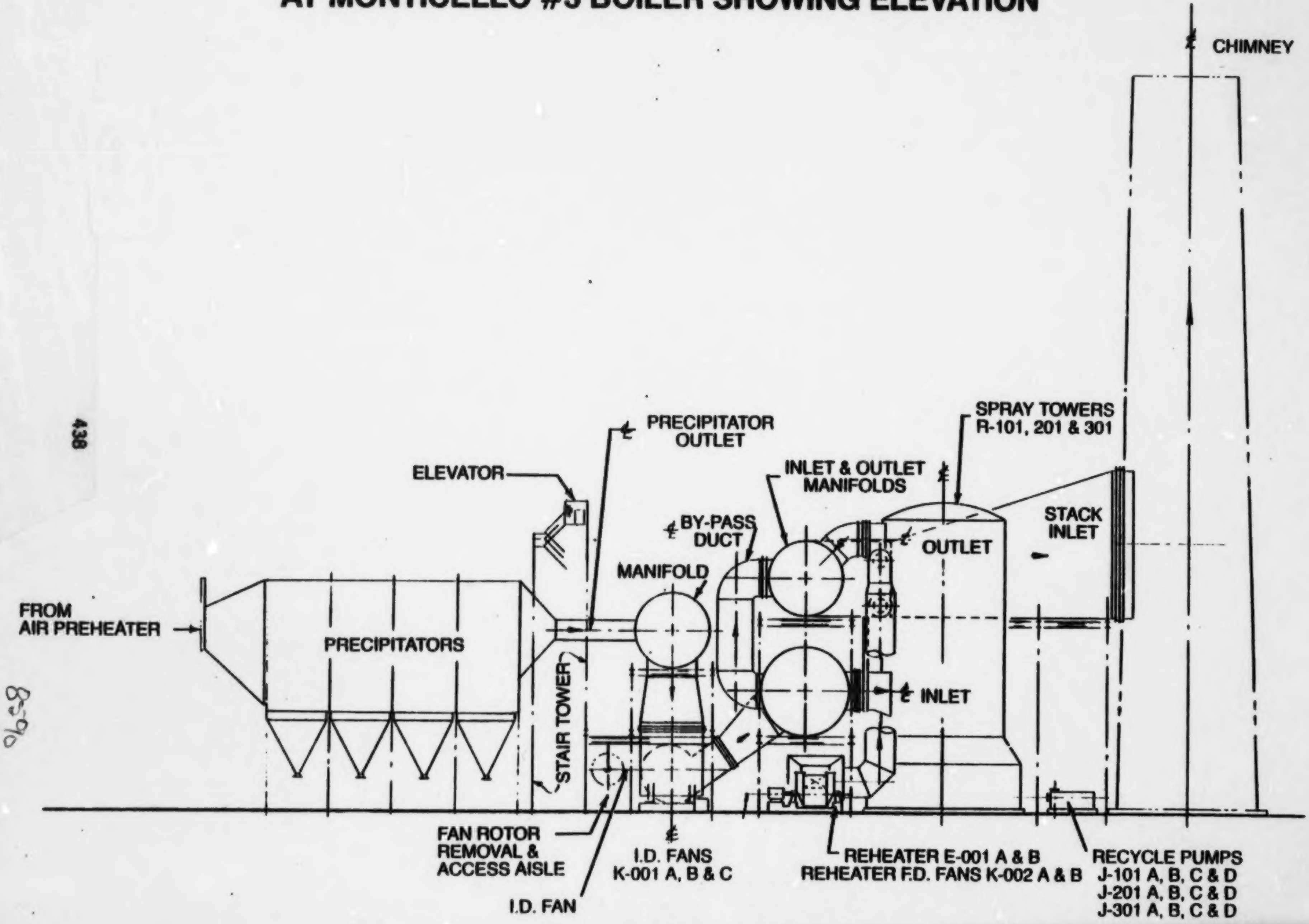
SIMPLIFIED FLOW DIAGRAM OF THE FGD SYSTEM AT MONTICELLO #3 BOILER OF TEXAS UTILITIES SERVICE


GAS OUT
TOP WASH SPRAYS
SERVICE WATER
BOTTOM WASH SPRAYS
MIST ELIMINATOR
LIMESTONE SLURRY FROM GRINDING SYSTEM
A
B
C
D
BLEED TO RETENTION POND
8"
6"
BLEED FROM PUMPS J-101 B, C, D
ABSORBER SPRAYS
TO TOWER R-301
TO TOWER R-301
RETENTION POND WATER MAKE-UP
6"
MAKE UP BY-PASS
6"
24"
GAS IN
POND WATER FOR DILUTION
6"
8"
4"
LIMESTONE SLURRY RECYCLE LOOP
BLEED RETURN BY PASS
12"
6"
SPRAY TOWER MAKE-UP
EMERGENCY OVERFLOW
2"
10"
AGITATORS
3"
24"
12"
J-101A RECYCLE PUMP
PUMP SUCTION
LIMESTONE SLURRY FEED TO R-101
PUMP SUCTION
SEAL POT
J-002A LIMESTONE SLURRY PUMP
G-003 LIMESTONE SLURRY STORAGE TANK
TO SPRAY HEADERS B, C, D
R-101 SPRAY TOWER

01688

FIGURE 2

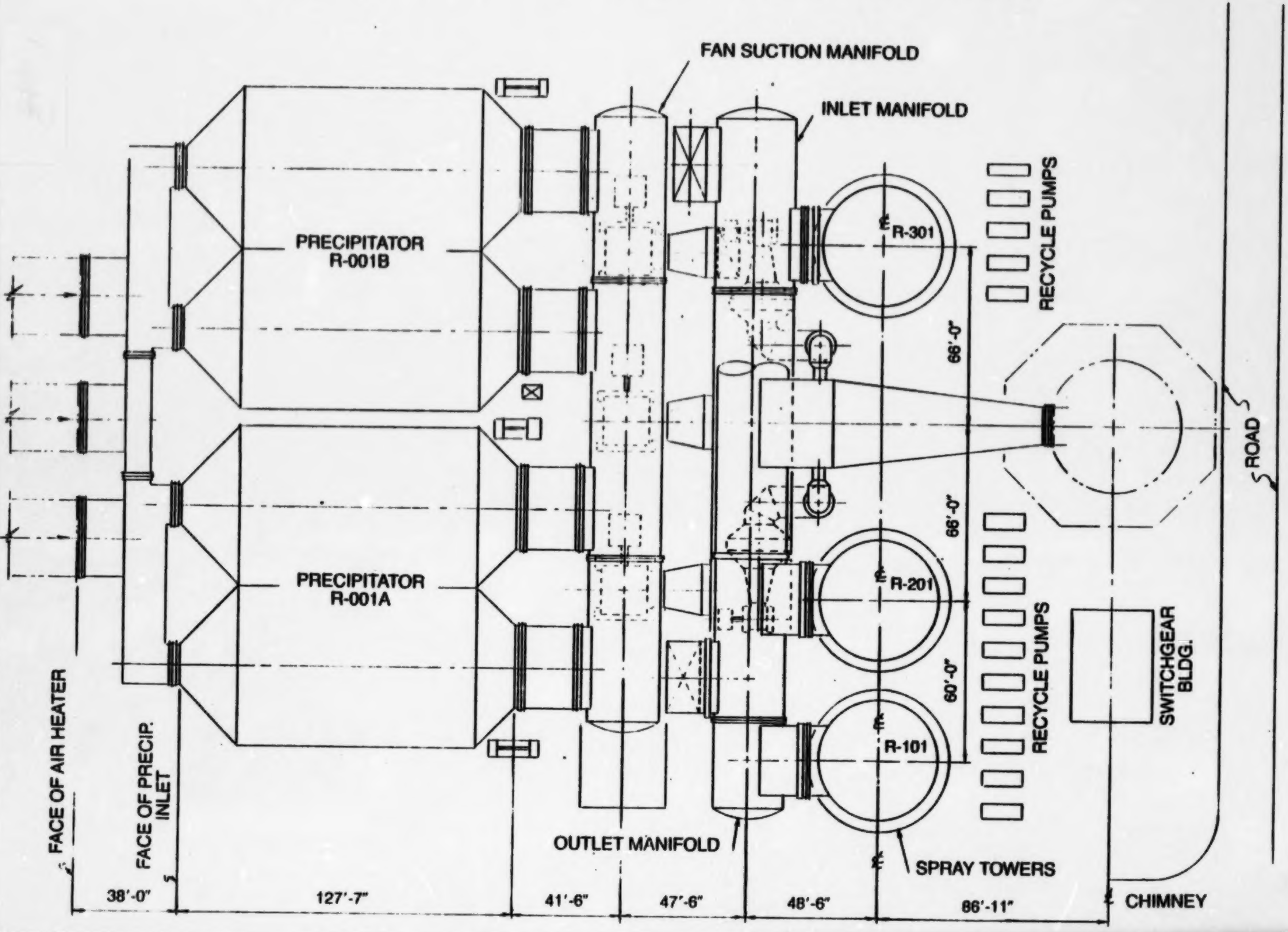
GENERAL ARRANGEMENT OF FGD SYSTEM AT MONTICELLO #3 BOILER SHOWING ELEVATION


CHIMNEY
SPRAY TOWERS
R-101, 201 & 301
PRECIPITATOR
OUTLET
ELEVATOR
INLET & OUTLET
MANIFOLDS
BY-PASS
DUCT
STACK
INLET
OUTLET
MANIFOLD
FROM
AIR PREHEATER
PRECIPITATORS
STAIR TOWER
INLET
FAN ROTOR
REMOVAL &
ACCESS AISLE
I.D. FANS
K-001 A, B & C
I.D. FAN
REHEATER E-001 A & B
REHEATER F.D. FANS K-002 A & B
RECYCLE PUMPS
J-101 A, B, C & D
J-201 A, B, C & D
J-301 A, B, C & D

8590

FIGURE 3

GENERAL ARRANGEMENT OF FGD SYSTEM AT MONTICELLO #3 BOILER SHOWING PLOT PLAN


FAN SUCTION MANIFOLD
INLET MANIFOLD
RECYCLE PUMPS
R-301
PRECIPITATOR
R-001B
66'-0"
ROAD
66'-0"
PRECIPITATOR
R-001A
R-201
RECYCLE PUMPS
SWITCHGEAR
BLDG.
60'-0"
R-101
FACE OF AIR HEATER
FACE OF PRECIP. INLET
OUTLET MANIFOLD
SPRAY TOWERS
38'-0"
127'-7"
41'-6"
47'-6"
48'-6"
86'-11"
CHIMNEY

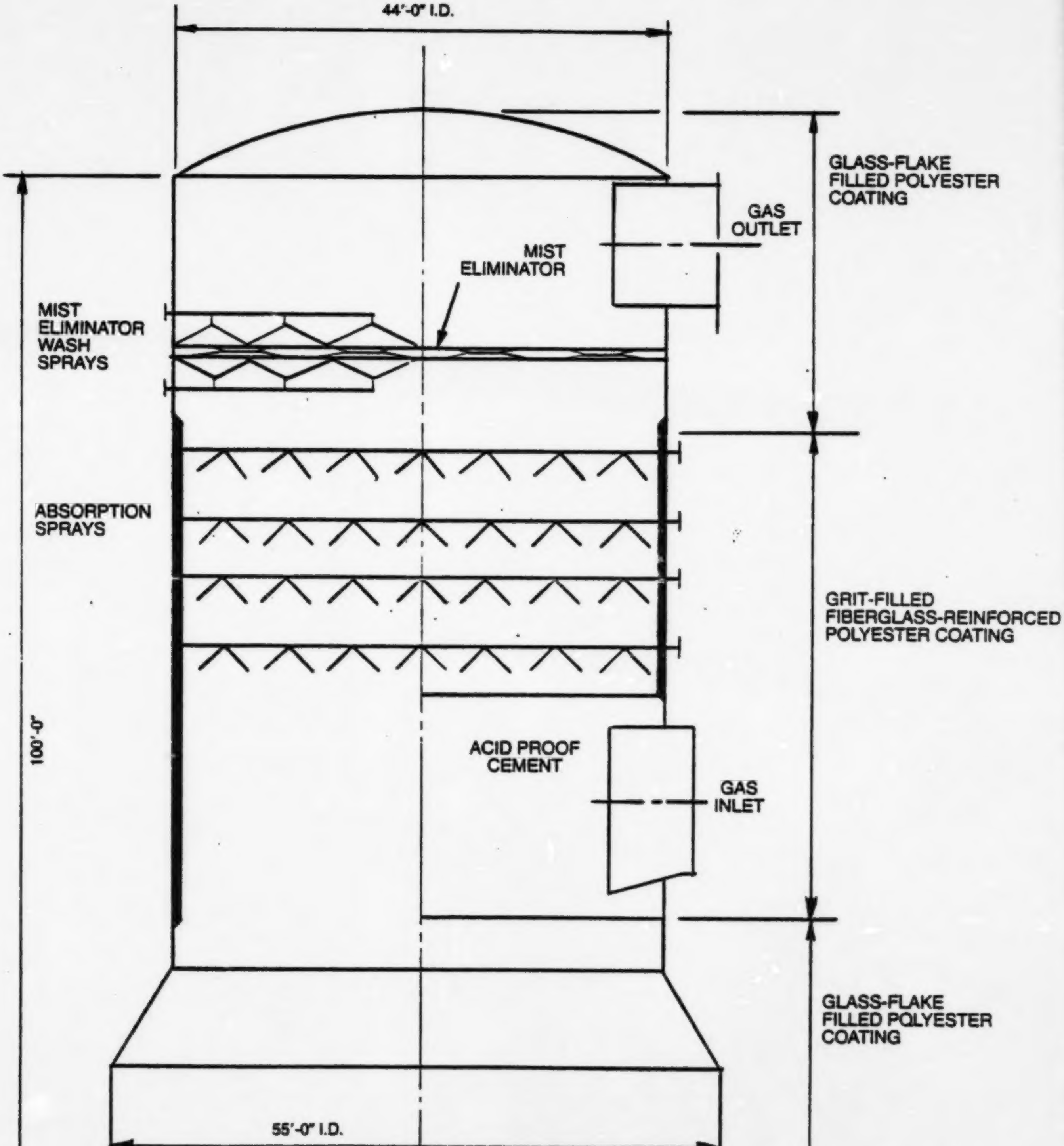
8596

is automatically controlled to maintain a predetermined SO_2 level in the stack. The control signal is provided by the SO_2 analyzer monitoring the stack gas. A supplemental reheat system is also provided for use when bypass gas is not sufficient to give the minimum superheat of 25°F. The supplemental reheat system consists of two parallel steam heat exchangers for heating the ambient air to 300°F which is injected into the outlet duct leading to the stack. The ambient air is driven by two centrifugal fans. Model tests for gas mixing were conducted to determine the location of bypass connecting ducts as well as point of injection of the supplemental steam heated air. The bypass gas is injected into breachings of the outlet manifold while the hot air is injected through four opposing ports in the outlet duct leading to the stack. (See Figures 2 and 3).

The SO_2 scrubbing is accomplished by three self-supporting spray towers with integral slurry recycle tanks. Each tower is equipped with a single blade guillotine damper on the inlet and a single louvre damper on the outlet. The single louvre damper can be used for flow balancing if required. A profile of the spray tower is shown in Figure 4. Each tower is 44 feet in diameter in the area of gas flow and expands to 55 feet to accomodate the recycle tank. The limestone slurry in the recycle tank is kept in suspension by four side mounted agitators. Each spray tower is equipped with four spray headers which are fed separately by dedicated centrifugal pumps of about 16,000 GPM nominal capacity. The slurry in each tower is sprayed through 200, 3 inch size hollow cone nozzles made of silicon carbide.

FIGURE 4

VERTICAL PROFILE OF THE SPRAY TOWER SHOWING MATERIALS OF CONSTRUCTION


44'-0" I.D.
GLASS-FLAKE FILLED POLYESTER COATING
GAS OUTLET
MIST ELIMINATOR
MIST ELIMINATOR WASH SPRAYS
ABSORPTION SPRAYS
GRIT-FILLED FIBERGLASS-REINFORCED POLYESTER COATING
100'-0"
ACID PROOF CEMENT
GAS INLET
GLASS-FLAKE FILLED POLYESTER COATING
55'-0" I.D.

Each spray tower has an integral four pass, open louvre vane mist eliminator which can be washed from both sides by spraying makeup water. The bottom side is continuously washed in sequence by actuating sprays in 12 pie shaped segments. The top is infrequently washed in a similar fashion as required. Prior to construction of the spray towers, gas distribution model tests had revealed that resistance imposed by the sprays was sufficient for uniform gas distribution, hence no additional gas distribution aids are installed.

Limestone quarry tailings are received by rail car and stored under covered shed. From the storage pile the limestone is conveyed into feeder hoppers for the wet ball mill grinders. One operating and one spare mill is provided, each with a capacity of about 30 tons per hour with product consistency of 90% minus 200 mesh. The ground limestone slurry is stored in a day tank and pumped to the spray towers as required for pH control in each operating spray tower. The pH is automatically controlled by signals from pH meters sensing the pH of slurry leaving the recycle tank.

The spent slurry bleed from each tower, taken under level control, is disposed of in an on-site pond. The reclaimed water from the dispsoal pond is recycled to the FGD system.

The density of the recirculating slurry is held at 8-10% solids by the use of nuclear type density analyzers, which control the amount of reclaimed water added to the towers.

MATERIALS OF CONSTRUCTION

All equipment and duct work upstream of the spray tower dealing with hot gas is carbon steel. Spray towers and the downstream duct work up to the stack are fabricated from carbon steel which is protected against corrosion and erosion by various linings. Each spray tower inlet duct starting from the guillotine damper and leading some distance into the tower is lined with acid proof cement (Prekrete G8). The base of the tower up to the liquid level is lined with glass filled polyester (Ceilcote 103). Above the liquid level and up to the top spray level is lined with 1/8 inch thick glass filled polyester lining (Ceilcrete 2500 AR) which incorporates special grit to provide abrasion resistance. The remaining portion of the tower above the top spray level up to the outlet isolation damper is glass filled polyester lined. The ductwork downstream of the outlet isolation damper, the outlet manifold and the duct leading up to the stack are lined with acid proof cement (Prekrete G8). The slurry recycle pumps and associated piping are rubber lined. The outside of the spray piping inside the tower is lined with abrasion resistant polyester lining (Ceilcrete 2500 AR). The spray nozzles are made from silicon carbide to provide abrasion proof service. The mist eliminator is constructed from fire retardent polypropylene and supported on FRP beams. The support beams for the slurry spray piping are rubber lined. The agitators for the recycle tank are also rubber lined.

The limestone slurry preparation section is carbon steel, except for the ball mills, recycle pumps and piping, cyclone classifiers and agitators which are rubber lined to guard against abrasion.

The dampers upstream of the spray towers including the bypass louvre are carbon steel construction. The expansion joints in this area are carbon steel and layered asbestos fabric construction. The dampers at the outlet of the tower are 316L stainless steel while the expansion joints are asbestos filled viton.

SYSTEM PERFORMANCE

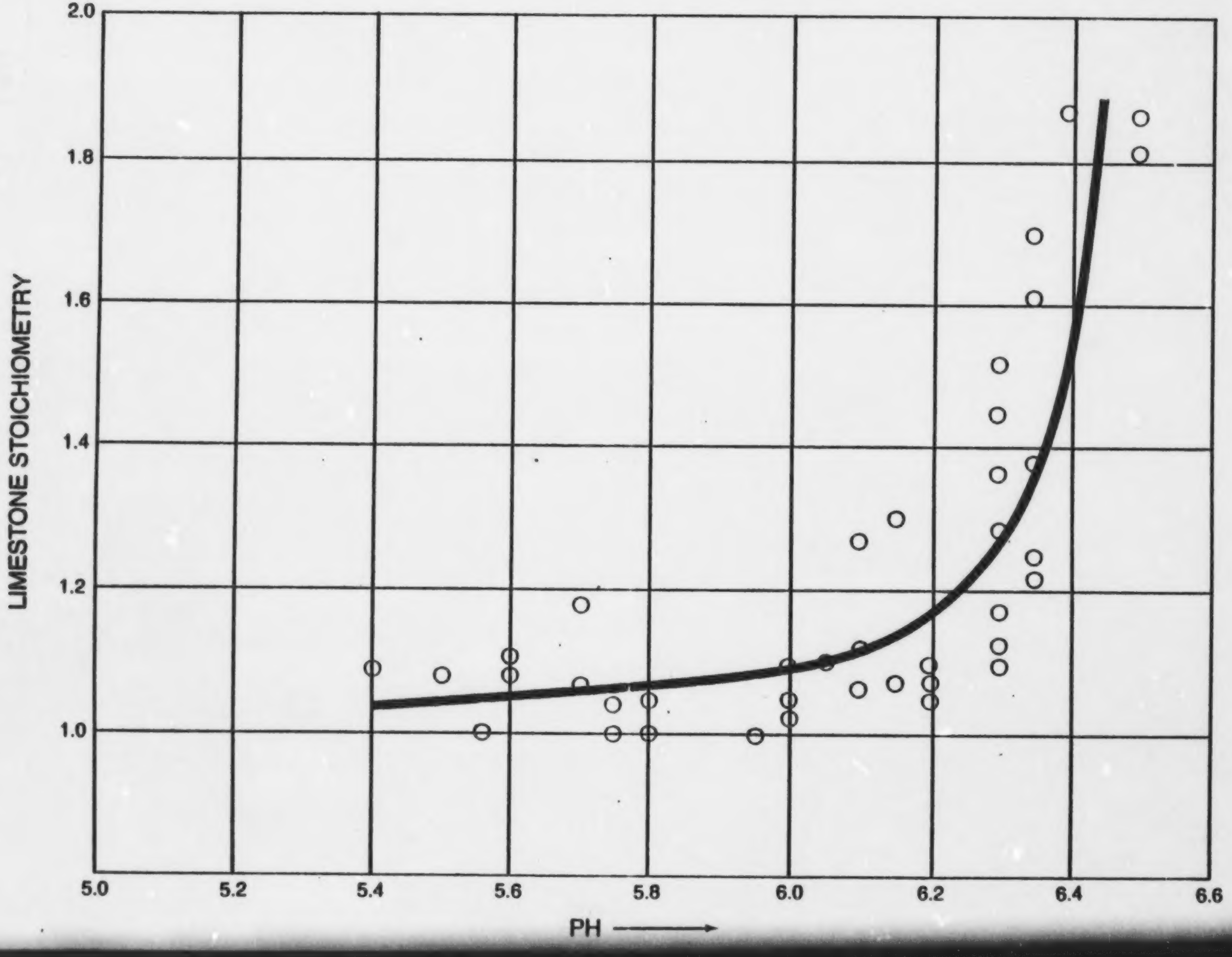
The FGD system was first placed in service on August 18, 1978. Since that time performance of the system has been good. The system easily met EPA compliance requirements during testing conducted during June, 1979. The SO_2 removal efficiency of the towers has generally been 95% or better. As a result of this efficiency, it has been possible to operate through the unit load range with two towers under most conditions; utilizing the third tower as a standby unit. However, due to the recycle pump situation, to be discussed later, and the desire to maintain the highest possible compliance with regulatory agency standards, it has been necessary to operate with all three towers in service for the last year. This has resulted in satisfactory operation even though two and often three of the towers have been operated with only two recycle pumps in service for extended periods.

During the operation of the scrubber to date the sulfur content of the fuel has varied from 0.4% to 1.7%, with a typical range of 0.6% to 0.8%. The E.E.I. availability of the scrubber has been 99.5% or greater. Texas Utilities Generating Company utilizes a "compliance factor" as a better indication of the true performance of an FGD system. The "compliance factor" is determined by dividing the number of hours of operation within SO_2 compliance limits by the total hours of generation. This information is shown in Figure 6. The compliance factor reflects non-compliance excursions resulting from all factors. The low compliance readings during the first few months of operation as indicated in Figure 6 resulted from problems outside of the FGD system primarily related to the precipitator performance. Figure 6 also shows the limitation factor which is a measure of generation loss due to FGD system. During

FIGURE 5

LIMESTONE UTILIZATION CURVE FOR THE MONTICELLO #3 FGD SYSTEM.

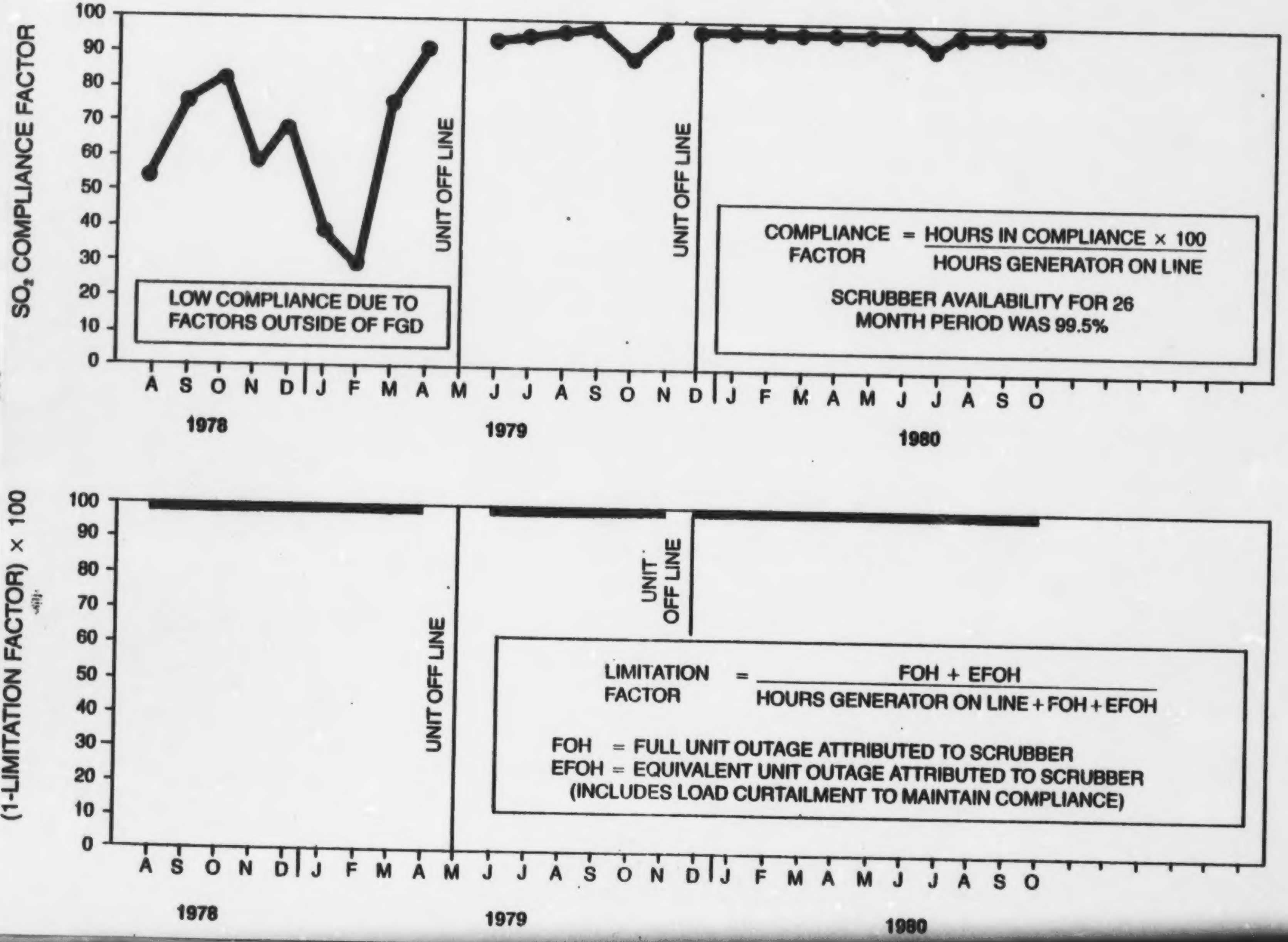
Stoichiometry defined as pound moles of limestone used per pound mole of SO_2 removed as a function of recirculating slurry pH.



80%

FIGURE 6

SO_2 SCRUBBER COMPLIANCE & LIMITATION FACTOR FOR MONTICELLO #3



the 26 month period shown here, the power generation loss attributable to the FGD system has only been a fraction of a percentage.

POWER REQUIREMENTS

Full load auxiliary power consumption of the SO_2 removal system including I.D. fans is approximately 10 MW. This is based on the assumption that 35% of the I.D. fan power consumption is due to the scrubber operation. When the limestone grinding system is in service, the auxiliary load is increased by 0.6 MW. The limestone grinding system has had a duty cycle of 5 - 6 hours per day when the unit is operating at near full load and the sulfur content of fuel is in the range of 0.7 - 0.9%.

REAGENT REQUIREMENTS

The limestone utilization is pH dependent as shown in Figure 5. When the system is operated within a pH range of 5.5 - 6 the limestone stoichiometry based on absorbed SO_2 is between 1 to 1.10.

MANPOWER REQUIREMENTS

The spray tower system has been relatively easy to operate and maintain, consequently the manpower requirement for operation has been modest. Since the flue gas controls are intergrated into the boiler train, the BTG operator can also control the flue gas flow to the spray towers.

The following is the list of personnel dedicated to scrubber operation:

System Area	Personnel Per Shift
Scrubber Control	1
Limestone Handling and Milling	1/2
Chemical Technician	1/3
Environment & Instrument Technician	1/4
Mechanical Maintenance	1 1/2
Electrical Maintenance	1/4
Total	3 5/6

In order to have 24 hour a day seven days a week coverage, a total manpower of 15-1/3 men is dedicated to the scrubber operation.

OPERATION EXPERIENCE

During the first four months of operation several breaks in the fiberglass line that supplies reclaim water from the sludge disposal pond to the towers were experienced due to poor make-up of joints during original installation and vibration due to inadequate support and restraint of piping in the area. This has been the only problem that resulted in the removal of a scrubber when the generator was on the line. The problem was corrected with the replacement of the fiberglass line with carbon steel pipe in the areas of failure.

Difficulty has been experienced in moving limestone from the storage pile to the grinding system due to pluggage of the hoppers and mechanical failure of the feeders and associated equipment. Pluggage is a problem due to the nature of the limestone used, which is a by-product of crushed limestone production and contains a high percentage of fine material and moisture. This results in pluggage in the reclaim hoppers especially when wet or when stacked high. It has been necessary to feed the reclaim hopper with a front end loader.

Several problems have been experienced with the tower inlet guillotine dampers. The bottom seals have been damaged due to ash and sludge accumulation in the seal trough. Several of the jack screws and pushrods have been damaged due to binding of the dampers. Minor linkage problems were experienced on the by-pass dampers; otherwise the tower outlet and by-pass dampers have performed well. Since no internal maintenance has been required while the scrubber has been in service, it is not known whether damper leakage would permit safe entry while on the line.

Several minor instrumentation problems have been experienced. Excessive drift has been a problem with the density control instrumentation. The 0-14 range of the pH instrumentation originally supplied was too wide to allow good control in the narrow range of 5.6 - 5.8. The scale was expanded and the pH system has performed satisfactorily. The system is easily operated manually with the result that instrumentation problems have not had any appreciable impact on the operation of the scrubber.

The rubber lining of the side mounted agitator blades has failed at the tips allowing erosion damage to all agitator blades.

The most significant problem experienced with the FGD system has been repeated failures of the rubber lining of the slurry recycle pumps. Although this problem has not resulted in the loss of availability of the FGD system or noncompliance with emission limits, it has resulted in a very significant maintenance expense. Efforts are continuing to resolve this problem by reducing the speed of the pump. A different manufacturers' pump has also been installed for testing.

Another problem was experienced when an attempt was made to use ash water on one tower for mist eliminator wash rather than fresh water. The high levels of calcium sulfate in this water resulted in extreme fouling of the mist eliminator packing material. The high velocity of gas through the unplugged areas combined with the increased load on the other towers resulted in slurry carryover into the outlet duct and chimney.

With the exception of the recycle pump problem, the operating experience with the system has been relatively good. There have been no lining problems in the towers. Spray nozzle plugging has not been a problem and no significant problems have been experienced with the internal piping other than external erosion due to impingment in isolated areas. It has not yet been necessary to remove a tower from service for internal maintenance.

OPERATING AND MAINTENANCE COSTS

Operating and maintenance costs are summarized in Table 2. Over the first 22 months of operation the FGD system operating labor cost has averaged $7,222 per month. The maintenance material and labor cost for the scrubber has averaged $65,396 per month with material only averaging at $35,341. The maintenance material and labor cost for the limestone preparation system has averaged $15,388 per month.

A significant portion (estimated at 40-50%) of the scrubber maintenance material and labor cost has been due to the recycle pump problems. Resolution of this one problem will significantly reduce maintenance cost.

TABLE 2

OPERATING AND MAINTENANCE COSTS FOR MONTICELLO #3 FGD SYSTEM

Calendar Month	Scrubber Operation	Maintenance Labor and Material	
		Scrubber Area	Limestone Area
A/78	$ 7,000*	$ 5,651	$ 3,598
S	7,000*	28,085	7,439
O	7,000*	10,066	5,312
N	7,000*	12,670	12,351
D	7,000*	27,307	12,376
J/79	7,855	15,149	10,881
F	7,855	24,007	6,703
M	7,855	30,730	15,587
A	7,855	32,506	9,988
M	4,309	78,383	3,821
J	8,028	24,597	11,866
J	6,863	44,128	19,933
A	8,117	60,128	19,147
S	6,890	51,222	6,625
O	7,088	44,225	17,610
N	7,763	127,713	26,351
D	9,695	158,096	7,836
J/80	7,521	49,484	12,199
F	7,891	119,228	45,223
M	6,695	70,494	36,621
A	7,110	294,694	15,873
M/80	4,488	130,144	30,185
Total	158,873	1,438,707	338,525
Monthly Average	7,222	65,396	15,388

*Estimated Costs

ADDITIONAL ESTIMATED MONTHLY COSTS AVERAGED OVER 22 MONTH PERIOD

Chemical Tecnician	1,000
Instrument Technician	500
Supervisory	500
Total Additional Costs	2,000

88%

DUAL ALKALI DEMONSTRATION PROJECT INTERIM REPORT

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ABSTRACT

This paper will discuss the results of the recently performed acceptance test on the dual alkali system serving Louisville Gas and Electric Company's Cane Run Unit 6 boiler. The acceptance test was conducted to measure the system performance with respect to the guarantees offered Louisville Gas and Electric by Combustion Equipment Associates. The results of the testing were as follows:

- SO_2 removal averaged 94% and 143 ppm outlet concentration
- Soda ash consumption averaged 0.042 mole soda ash per mole sulfur dioxide removed
- Lime consumption averaged 1.04 mole CaO per mole sulfur dioxide removed
- Power consumption averaged 1.05% of generation
- Filter cake solids averaged 52.2 wt % insoluble solids
- There was no net particulate matter addition

Various problems attributable to the boiler, the FGD system, and the quality and quantity of the carbide lime supplied to the system delayed the acceptance testing until July 1980. The year-long demonstration period was officially started in May 1980. The nature of the problems experienced and their solutions are discussed.

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NOTES

1. Company Names and Products

The mention of company names or products is not to be considered an endorsement or recommendation for use by the U.S. Environmental Protection Agency.

2. Units of Measure

EPA policy is to express all measurements in Agency documents in metric units. When implementing this practice will result in undue cost or difficulty in clarity, IERL-RTP provides conversion factors for the non-metric units. Generally, this paper uses British units of measure.

The following equivalents can be used for conversion to the Metric System:

British	Metric
5/9 (°F-32)	°C
1 ft	0.3048 m
1 ft^2	0.0929 m^2
1 ft^3	0.0283 m^3
1 grain	0.0648 gram
1 in.	2.54 cm
1 $in.^2$	6.452 cm^2
1 $in.^3$	16.39 cm^3
1 lb (avoir.)	0.4536 kg
1 ton (long)	1.0160 m tons
1 ton (short)	0.9072 m tons
1 gal.	3.7854 liters

DUAL ALKALI DEMONSTRATION PROJECT INTERIM REPORT

INTRODUCTION

The Dual Alkali Demonstration Project is a joint effort by a number of organizations under the sponsorship of the Environmental Protection Agency. The process being demonstrated is a sodium based concentrated mode using carbide lime as a regenerant. Louisville Gas and Electric Company (LG&E) is the owner-operator of the dual alkali system serving their Cane Run Unit 6 boiler, which is a nominal 280 MW high-sulfur coal-fired boiler (3.5-4.0% S). The design was developed by Combustion Equipment Associates (CEA) and Arthur D. Little, Inc. (ADL). The system was erected by the construction department of LG&E under the guidance of CEA/ADL at total cost of about $22 million (1976-1980 dollars) or about $79 per kW installed generating capacity (including waste disposal).

A process flow schematic of the dual alkali process at Cane Run 6 is depicted in Figure 1. Flue gas from the boiler passes through the electrostatic precipitators and is fed to two absorbers. A recycling sodium sulfite solution, flowing countercurrent to the flue gas across two stainless steel perforated plate trays, absorbs SO according to the following reaction:

1 $SO_3^{=} + SO_2 + H_2O \rightarrow 2HSO_3^{-}$

In addition, due to the absorption of sulfur trioxide from the gas and due to the oxidation of sulfite ion in solution, sulfate ($SO_4^{=}$) is formed in the absorbent liquor:

2 $H_2O + SO_3 \rightarrow H_2SO_4 \rightarrow 2H^{+} + SO_4^{=}$

3 $SO_3^{=} + 1/2\ O_2 \rightarrow SO_4^{=}$

The scrubbed flue gas is reheated by combustion gases from a direct oil-fired reheater and is ducted to the stack.

Sodium carbonate is added to either the thickener or the absorber to make up for losses of sodium in the system. Bleed streams of the spent aborbent solution from the absorbers are sent to the regenerator reactor trains where carbide lime is added to convert the bisulfite (HSO_3^{-}) in the spent absorbent, to sulfite ($SO_3^{=}$) in the regenerated absorbent, precipitating a mixture of calcium sulfite and sulfate solids:

4 $2HSO_3^{-} + Ca(OH)_2 \rightarrow CaSO_3\downarrow + SO_3^{=} + 2H_2O$

5 $SO_4^{=} + 2HSO_3^{-} + Ca(OH)_2 \rightarrow CaSO_4\downarrow + 2H_2O + 2\ SO_3^{=}$

The mixed solids actually can be designated as: $x\ CaSO_3 \cdot y\ CaSO_4 \cdot z\ H_2O$ where the ratio x:y is usually greater than 4 and z represents

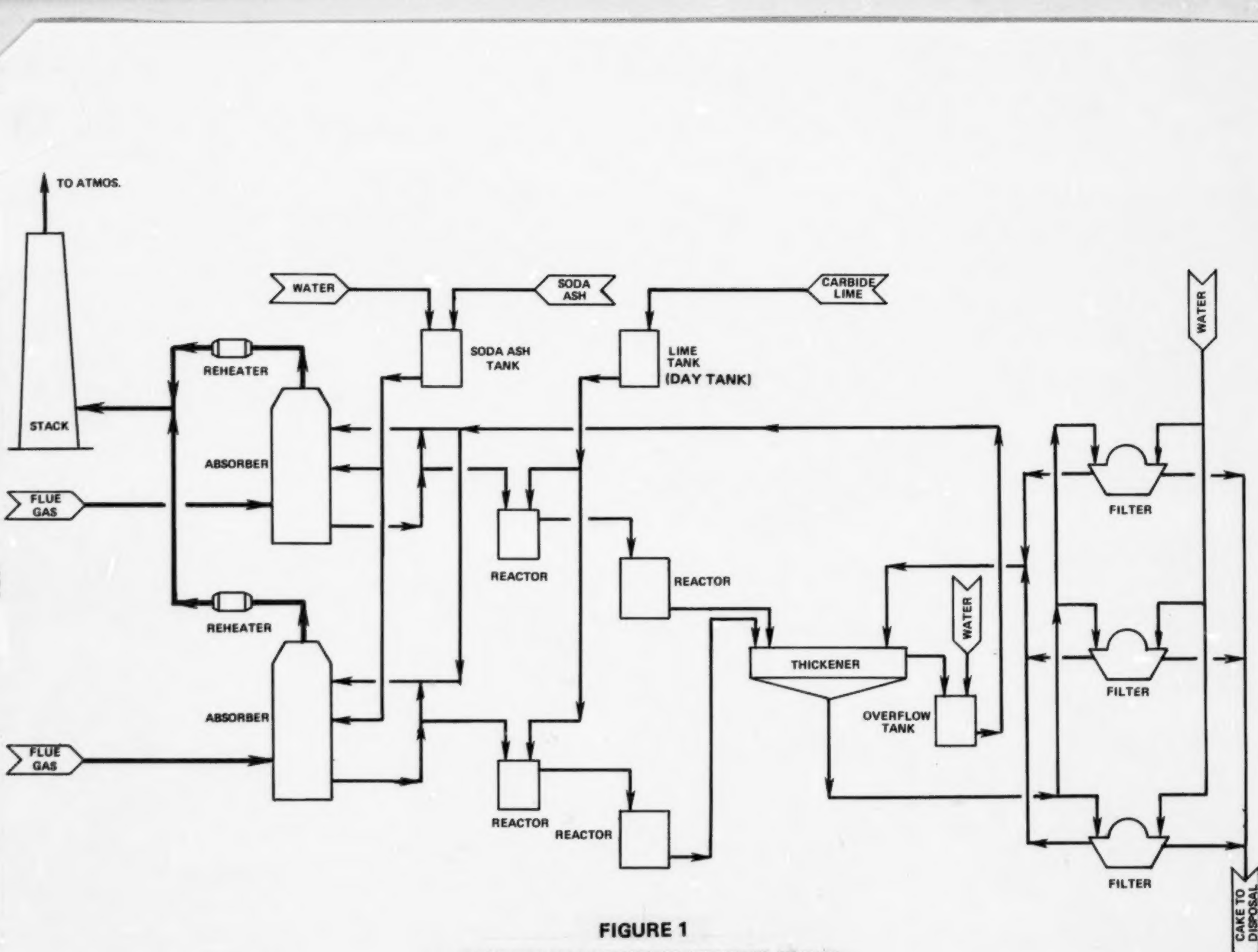
TO ATMOS.
STACK
WATER
SODA ASH
CARBIDE LIME
WATER
REHEATER
SODA ASH TANK
LIME TANK (DAY TANK)
ABSORBER
FLUE GAS
REACTOR
REACTOR
FILTER
WATER
THICKENER
FILTER
REHEATER
ABSORBER
OVERFLOW TANK
FLUE GAS
REACTOR
REACTOR
FILTER
CAKE TO DISPOSAL

FIGURE 1
DUAL ALKALI SYSTEM AT CANE RUN 6

some amount of water of hydration. No pure gypsum phase is formed. The solids are separated from the liquor in a thickener and are removed from the system on washed vacuum filters. The filter cake is mixed with fly ash and quicklime in a system designed by I.U. Conversion Systems. After fixation the solids are trucked to a landfill site for disposal. The clear liquor overflowing from the thickener is returned to the absorber recycle loop.

A comparison between the design basis and observed operation is given in Table 1. The design basis is taken from the design manual produced under this project, one of the sources of information to which the reader is referred for additional detail (References 1, 2, and 3). The system is designed to operate with a liquid to gas ratio of less than 10 gal./10^3 acf including liquor feed to the tray and spray recycle (typical lime or limestone slurry process are designed for about 50 gal./10^3 acf). The design flue gas pressure drop from the booster fan to the stack entrance is 8.5 in. of water.

Bechtel National, Inc. is under a separate contract with EPA to provide an independent test program to assess the operation of the system with regard to its performance guarantees, and to provide a demonstration program designed to characterize the system and monitor its performance over a year-long demonstration period.

Construction was completed in March 1979 and the system was initially charged and started up in April 1979. Various problems attributable to the boiler, the FGD system, and the quality and quantity of lime supplied to the system delayed the acceptance testing until July 1980. The year-long demonstration period officially started in May 1980. The problems and solutions are discussed later.

The acceptance test was conducted from July 17 to July 28, 1980. With one minor exception (filter cake quality), the system proved to be capable of successfully meeting its performance guarantees.

ACCEPTANCE TEST RESULTS

The 12-day acceptance test was conducted to measure the performance of the dual alkali system with respect to the guarantees provided to Louisville Gas and Electric Company by Combustion Equipment Associates.

Seven guarantees concern the operation in the following areas:

- Sulfur dioxide removal
- Carbide lime consumption
- Soda ash consumption
- Particulate matter emissions
- Power consumption
- Filter cake quality
- Year-long system availability

TABLE 1

Performance Conditions

	Design	Observed
Coal (Dry Basis)		
Sulfur	5.0% S	3.7% S (ave.)
Chloride	0.04% Cl	0.02% Cl (ave.)
Heat Content	11,000 Btu/lb	10,650 Btu/lb (ave.)
Inlet Gas:		
Flow Rate (Volumetric)	1,065,000 acfm	1,045,000 acfm (max.)
Temperature	300°F	280°F (max.)
SO_2	3471 ppm	2323 ppm (ave.)
O_2	5.7%	6.7% (ave.)
Particulate	0.10 lb/10^6 Btu	0.84 lb/10^6 Btu (ave.)
Outlet Gas:		
SO_2	$\leq$200 ppm	143 ppm (ave.)
Particulate	0.10 lb/10^6 Btu	0.10 lb/10^6 Btu (ave.)
Boiler Operation:		
Generation	280 MW	240 MW (max.)

Table 2 summarizes the guarantees offered and the corresponding results of the acceptance test. A brief discussion of each of the guarantee tests performed during the acceptance test follows.

Sulfur Dioxide Removal

The primary method of determining SO_2 removal relied on the continuous Lear Siegler monitor installed in the stack. This analyzer was certified in December 1979 by an outside contractor according to the procedure specified in the Federal Register. During the acceptance test, as a backup to the continous monitor and as an ongoing confirmation of the analyzer accuracy, wet chemical tests of the stack effluent according to EPA Method 6 were also performed daily, in conjunction with the particulate tests.

Preliminary results from the wet chemical analysis showed a discrepancy between these measurements and the continuous monitor readout. After an extensive check of the system, a burned ground wire was discovered in the signal line of the Lear Siegler continuous SO_2 monitor. From the data on the calibration sequences of the analyzer prior to, during, and after elimination of the grounding problem, it was concluded that the signal from the analyzer was offset on the low side by 30 ppm by the malfunction. Therefore the continuous stack SO_2 monitor readings for the first 7 days of the tests were corrected by 30 ppm. With this correction applied to the early readings, and subsequent to the repairs to the ground in the analyzer, the two techniques were in good agreement.

Both measurements showed that the system could meet the 200 ppm SO_2 outlet concentration guarantee. Table 3 summarizes the 24-hour average SO_2 results for the 12-day acceptance test. Table 4 summarizes the simultaneous wet chemical and continuous monitor measurements (the Method 6 tests were conducted only for the first 10 days).

Lime Consumption Guarantee

The lime consumption guarantee was specified as "not [to] exceed 1.05 moles of available CaO in the lime per mole of SO_2 removed from the flue gas". Lime consumption was determined by analyzing representative samples of filter cake collected as the cake was discharged from the filters prior to fixation. The cake was analyzed for total calcium and total sulfur. The total calcium represented the lime used, and the total sulfur represented SO_2 removed from the flue gas. A portion of the calcium entering the system with the carbide lime is present as carbonate and therefore does not represent alkalinity available for regeneration. Each time the lime day tank was filled, a sample of lime was analyzed for available alkalinity and total calcium. From these results, a correction factor was developed to account for unreactive calcium in the carbide lime feed. During the 12-day acceptance test the calcium consumption, corrected for available alkalinity as described above, averaged 1.04 moles of available CaO per mole of SO_2 removed, thus meeting the guarantee which required less than 1.05 moles/mole of sulfur removed. Table 5 summarizes the analyses performed on the filter cake samples.

TABLE 2

Performance Guarantees and Acceptance Test Results

GUARANTEE	TEST RESULTS
SO_2 Removal	
200 ppm dry basis (D.B.) without additional air dilution	143 ppm (D.B.) without additional air dilution
Calcium Consumption	
1.05 moles available CaO per mole SO_2 removed	1.04 moles available CaO per mole SO_2 removed
Soda Ash Consumption	
0.045 moles Na_2CO_3 per mole SO_2 removed	0.042 moles Na_2CO_3 per mole SO_2 removed
Net Particulate Addition	
No net particulate addition by FGD system	Net particulate removal averaging 88% efficiency
Power Consumption	
System will consume (excluding reheat) not more than 1.2% of power generated at peak capacity	System consumed 1.05% of power generated
Filter Cake Properties	
Filter cake will contain a minimum of 55 wt % insoluble solids	Filter cake averaged 52.2 wt % insoluble solids

TABLE 3

Acceptance Test Continuous SO_2 Analysis

Acceptance Test Day	24 Hour Continuous SO_2 Analyzer Results (ppm, dry basis)			
	A Inlet	B Inlet	Stack	% Removal
1	2444	2418	130	94.7
2	2674	2570	129	95.0
3	*	2390	130	94.6
4	*	2290	152	93.4
5	2265	2315	157	93.1
6	2567	2515	140	94.5
7	2113	2021	132	93.6
8	2116	2088	124	94.1
9	2395	2339	146	93.8
10	2372	2315	171	92.7
11	2292	2233	156	93.1
12	2167	2166	130	94.0
Average	2340	2305	141	93.9

* Analyzer printout malfunction

TABLE 4

Acceptance Test Continuous Monitor and EPA Method 6 Analysis

Acceptance Day	Hours	SO_2 Concentration, ppm, dry basis					
		A Inlet		B Inlet		Stack	
		DuPont Analyzer	Method 6	DuPont Analyzer	Method 6	LSI Analyzer*	Method 6
1	1400-1700	2434	2330	2516	2330	119	124
2	1100-1300	2434	2150	2423	2180	122	163
3	1000-1300	2592	2210	2670	2290	117	154
4	1200-1500	2836	2390	2674	2480	155	159
5	1000-1300	2656	2330	2606	2410	136	137
6	1000-1300	2716	2350	2418	2480	184	212
7	1600-1900	2337	2040	2250	2030	**	137
8	1100-1300	2395	2120	2330	2100	113	130
9	1100-1400	2864	2530	2721	2450	122	133
10	0900-1200	2690	2410	2624	2360	160	137

* Analyzer readings for days 1-6 corrected for the effect of the burned out ground wire

** Analyzer out of service for repairs to ground wire

TABLE 5

Acceptance Test Daily Average Filter Cake Analysis

Test Day	As Received Basis			Insoluble Solids wt %	$\frac{\text{Mole } Na_2CO_3}{\text{Mole } SO_2}$	$\frac{\text{Mole } CaO}{\text{Mole } SO_2}$
	Na wt %	Ca wt %	Total Sulfur wt %			
1	0.55	14.88	31.35	52.65	0.037	1.139
2	0.58	14.60	31.58	52.20	0.038	1.109
3	0.70	15.64	32.60	52.60	0.045	1.151
4	0.48	15.72	32.68	53.72	0.031	1.154
5	0.45	15.15	31.92	53.92	0.029	1.139
6	1.11	15.20	33.20	51.90	0.070	1.099
7	0.58	14.35	31.80	50.70	0.038	1.083
8	0.50	14.44	32.32	51.40	0.032	1.072
9	1.07	14.05	32.58	51.00	0.068	1.035
10	0.45	14.43	33.07	52.43	0.028	1.047
11	0.77	14.85	33.48	52.82	0.047	1.064
12	0.62	13.74	31.64	50.58	0.041	1.042
Average				52.16	0.042	1.095

Calcium consumption corrected for available alkalinity (1.095 x 0.95* = 1.040)

* Correction factor developed from analysis of incoming carbide lime for mole of available alkalinity per mole of total calcium

Soda Ash Consumption

The soda ash consumption was determined by analysis of total sodium and total sulfur in the filter cake. According to this analysis the consumption of soda ash averaged 0.042 moles of Na_2CO_3 per mole of sulfur dioxide removed and therefore met the guarantee requirement of 0.045.

Particulate Matter Emission

The system was guaranteed not to make any net addition of particulate matter to the gas stream prior to discharge. Particulate tests, following EPA Method 5, were conducted on the inlets to the absorber modules and in the stack (downstream of reheaters) during the acceptance test. The results of 10 simultaneous tests showed convincingly that there was no net addition of particulate matter across the system. Actually, the absorber performed as a particulate removal device averaging 88% net removal of incoming particulate. Table 6 displays the results of particulate matter tests performed during the test program.

Although the FGD system met the guarantee requirements, the test was not very stringent due to the low level of performance by the electrostatic precipitator during the acceptance test period. The FGD system was originally designed to process an incoming flue gas stream containing the equivalent of 0.1 lb of particulate matter/10^6 Btu or less. During the acceptance test, however, the level of incoming particulate matter was almost an order of magnitude higher. Thus it is not surprising that the absorbers functioned to remove particulate matter even at the relatively low pressure drop at which they operated. The particulate matter emissions from the stack, however, were on the order of 0.1 lb/10^6 Btu as required for the Cane Run Unit 6 FGD system under the appropriate requirements to control particulate matter emissions.

Power Consumption

The system, excluding reheat, was guaranteed not to use more than 1.2% of the total power generated by the boiler/turbine unit at gross peak load. During the acceptance test the peak generation was 240 megawatts (MW). Correspondingly, the power consumed during peak generation was 2.5 MW, or 1.05%. The guarantee was met based on peak generation and also based on average generation over the the test period. During the 12-day test, the average load was 178 MW and the average power consumption by the FGD system was 2.05 MW, or 1.15%.

Waste Filter Cake Properties

The system was guaranteed to produce a waste filter cake containing a minimum of 55 wt % insoluble solids. The filter cake averaged 52.2 wt % insoluble solids during the acceptance test. While this fell slightly

TABLE 6

Acceptance Test Particulate Test Results

Acceptance Test Day	Particulate ($lb/10^6$ Btu)			
	A Inlet	B Inlet	Stack	% Removal
1	0.5320	0.7120	0.0895	85.6
2	0.6590	0.3620	0.0932	81.7
3	0.9470	1.0700	0.1110	89.0
4	0.9440	0.8060	0.1030	88.2
5	1.1100	0.9200	0.1020	90.0
6	0.9900	1.4900	0.1020	91.8
7	0.5890	0.8470	0.1020	85.6
8	0.6250	0.6490	0.0893	86.0
9	0.7890	1.2000	0.1100	88.9
10	0.9620	0.5930	0.1020	86.9
Average	0.8147	0.8649	0.1004	88.0

short of guarantee, the product discharged to the IUCS process was uniform in moisture content and was suitable for working into a stable and manageable product through the fixative process. Optimization of filter cloth selection and filter cycle will continue with the goal of showing that compliance with this guarantee can be met during the demonstration year.

System Availability

System availability, as defined by the Edison Electric Institute (available hours divided by the total hours in the period under consideration), was guaranteed to be greater than 90% for the demonstration year. While it is too early to report such a figure, through the first 4 months of the demonstration year (May-August), the availability of the system has averaged 99.8%.

OPERATING AND MAINTENANCE PROBLEMS

Up to the time of the acceptance testing there were a number of mechanical problems and a few chemical problems which affected system performance and led to cumulative delays in executing the program. None of the problems have been insurmountable, but their solutions have been time consuming. It is important to report the nature of these obstacles so that future installations of this or similar technology can benefit from the experience.

Recycle and Thickener Return Pumps

There have been two major problems with the high-capacity low-speed pumps for recirculation of absorbent liquor to the trays, and return of thickener overflow liquor to the absorbers. The first problem was the mechanical shearing of the impellers at the hub. The original pump impellers were manufactured in two parts: a body and a separate hub for attachment to the shaft. The hub was welded to the body. All of the impeller failures were on this welded seam. This problem was eliminated when the pump vendor supplied a one-piece molded impeller body.

The second major problem involved the rapid failure of the suction side of the pump liner. As a result of close tolerances between the casing liner and the impeller, the two surfaces were rubbing; the resulting abrasion destroyed the liner. After completely dismantling the pumps, it was discovered that a finishing step appeared to have been omitted at the factory, leaving about 1/4-in. excess length on each shaft. Milling each shaft to its design size eliminated this problem.

Mist Eliminator Collapse

Within a few months the startup, both absorber modules experienced high pressure drop problems. Inspection of the internal structure revealed that the mist eliminator sections had sagged or collapsed structurally. The problem was solved by replacing the mist eliminator sections with those of a different manufacturer. Since the replacement, in August 1979, there has been no further problem with the mist eliminators.

Tray Pluggage

One of the most perplexing problems was the pluggage of the absorber trays due to deposition. At first the observed deposit was thought to be carbonate scale resulting from pH upsets in the modules. Careful analysis showed the precipitate to be an aluminum-hydroxy-silicate complex. The mechanism of dissolution and subsequent deposition was traced to the operating pH of the reaction train. Aluminum was found to be entering the system with the carbide lime. At the operating pH in the reactor, above 11.5, the aluminum compound is soluble in the liquor. When the thickener overflow recycle combined with the recirculating absorbent, the resultant drop in pH caused the aluminum to precipitate on the absorber trays.

Reducing the operating pH of the reactor to between 10.0 and 10.8 reduced the solubility of the aluminum within the reactor and thickener. This change ahead of the absorber minimized the pluggage problem. At the reduced pH set point, however, there is less buffering and control of reactor pH is more difficult.

Water Balance

The system initially experienced a severe water imbalance. This was partly due to a lack of familiarity with the system, and partly because of low-solids concentration in the carbide lime feed. The other lime slurry systems at Cane Run can tolerate an occasional open-loop excursion. However, the dual alkali process must operate in a closed-loop at all times, since the high concentration of solubles in the scrubbing liquor makes disposal unacceptable for both environmental and economic reasons.

The system was designed to accommodate 70% water (30% solids) in the incoming carbide lime slurry. Initially the water content was consistently in the 82-85% range. At this concentration the system was receiving twice the design input water flow. After only a few hours of operation the volume of water in the system had accumulated to the point where the lime feed had to be cut off. The absorbers continued to function as evaporators until the water level dropped low enough to resume normal operation.

Strict control of the incoming lime concentration from the supplier and the addition of a ball mill-hydroclone system to remove oversize particles alleviated the problem.

Soda Ash Silo Pluggage

Soda ash is added to the system by a dry weigh feeder which feeds dry solids from a storage silo to a mix tank where it is mixed with absorbent liquor. Vented moisture vapor from the hot mix tank backs up into the weigh feeder screw conveyor and causes the soda ash to form lumps which prevent the smooth flow of feed to the system. The system had a small fan to blow the moisture-laden air back into the mix tank; however, it proved to be under-designed. Although a larger fan was installed to improve the situation, the soda ash feed system still remains a relatively high maintenance item.

Thickener Blockage

In mid-January of 1980, the thickener rake seized during a boiler outage for repair and ultimately required a shutdown and major overhaul of the thickener. This did not occur during normal operation, but rather during the transient period in which the boiler and FGD were being shut down for maintenance. The stoppage was postulated to have resulted from an overloading of the thickener with washings of accumulated solids (including fly ash) from the bottom of the absorber. Lacking a bottom drawoff, the absorber allowed fly ash to be trapped and accumulated in its lower portion. The problem could apparently have been avoided if the solids from the bottom of the absorber had been slowly pumped to the thickener while the thickener and filters continued in operation until the absorber bottom was purged of solids.

Correction of the problem took about 3 weeks, during which about 2 million gallons of liquid and solids had to be removed from the thickener (liquid was temporarily stored, and solids were impounded off site). To accomplish this, large access entrances were cut in the thickener sides to allow entry by personnel and equipment to dig out the compacted solids.

Overloading of the thickener has not recurred. The solids in the bottom of the absorbers are still not subjected to mechanical agitation, but they are no longer washed into the thickener in large slugs.

Sulfur Dioxide Monitoring

Sulfur dioxide measurement in the inlet to and the outlet from the absorbers is performed by continuous DuPont UV Model 460 SO_2 analyzers. In the stack, sulfur dioxide concentration of the scrubbed gas is measured by a Lear Siegler SO_2 analyzer.

Three problems have occurred in the measurement of SO_2 using the DuPont analyzers supplied with the dual alkali system:

- Plugging of the sample probe
- Maintaining a steady calibration of the instruments
- Stratification of scrubbed gas across the absorber exit duct

The first two problems have been minimized by daily inspections to determine if calibration or cleaning of the probes is required. An attempt to alleviate the last problem will be made by moving the SO_2 probes downstream of the reheaters, which should also help reduce the first two problems.

Failure of FRP Piping

The FRP (fiberglass reinforced plastic) piping in slurry service (i.e., thickener underflow and filter feed) has been a major maintenance item. Some failures have been spectacular, some minor. Late in the fall of 1979 a flush connection on the underflow line snapped off. Slurry from the thickener flooded the access tunnel below the thickener before the break could be isolated. Routinely, elbows in the line from the thickener to the filter have required repairs because of erosion damage and failure of the connection bond. Gradually all the underflow FRP piping is being replaced with mild steel. While mild steel has a limited life span in this service, failures will be less catastrophic.

pH Control

Reliable and accurate pH measurements for pH control in the reactors and in the scrubber bleed stream have been particularly bothersome. The pH related problems are attributed to:

- Inability to keep the probes clean
- Poor responsiveness of the probes
- Pluggage of the sample lines
- Poor calibration techniques

Experimentation with different instrument designs and sampling methods is gradually alleviating the first three problems. Detailed calibration instructions and cross checking of the results by two operating departments have minimized the last. On-line pH readings are compared daily with pH measurements taken with a portable pH meter by the LG&E scrubber laboratory personnel. If these readings are in disharmony by more than 0.3 pH units, the on-line probes are recalibrated.

All the original L&N pH probes have been replaced with Great Lakes models. To measure the pH of the primary and secondary reactors, a Great Lakes Model 60 submersible probe is placed in the overflow chute from primary to secondary reactor, and in the secondary reactor below the liquid level near the overflow. The pH of the bleed and thickener return streams is measured by Great Lakes Model 60 flow-through pH probes with ultrasonic cleaners.

Filter Operation

There have been two major concerns with the rotary vacuum drum filters. First, the cake quality has varied between 45% and 55% solids. Second, it has not always been possible to properly wash the cake to meet sodium consumption guarantee.

Prior to the acceptance test, experimentation with different filter cloths led to installation of a new filter cloth. The original cloth was a polypropylene cloth supplied by National Filter Media of Hamden, CN. During the acceptance test this cloth was replaced with a multifilament nylon cloth supplied by Thoerner Products Corp., of Pittsburgh, PA. The new cloth produced a more consistent quality cake but had a tendency to blind. During the acceptance test, cake washing was sufficient to meet the soda ash consumption guarantee, but the blinding detrimentally affected the percent solids of the filter cake. There is also some concern that poorer quality solids may be produced in the reactors at the lower pH levels required to control dissolution of aluminum and silicon compounds.

Proper cake washing on the filters is subject to a number of considerations. The wash water rate (as limited by the water balance), the quality of solids produced, the thickness of the cake (controlled by drum angular velocity), the wash spray configuration, and the quality of the filter cloth (blinding characteristics) are all important parameters. Therefore, experimentation with different filter cloths and varying operating parameters is continuing.

DEMONSTRATION PROGRAM PLANS

Although the results reported here have focused on startup and acceptance testing, it seems appropriate to outline the major work underway and planned as part of the demonstration program.

Commercial Grade Lime Testing

A month-long test, using commercial-grade lime in place of carbide lime, will be conducted as part of the demonstration program to confirm the interchangeability of the two materials for use in a lime dual alkali system. The carbide lime contains silicon and aluminum compounds that are potentially detrimental to the operation of the system, as previously noted. Bench-scale tests have already shown that commercial-grade lime is more reactive than carbide lime, and further improvement in lime consumption is expected during this test. Conversely, carbide lime is thought to contain an oxidation inhibitor not present in commercial lime. Much of the success of this system relies on the process liquor remaining subsaturated in calcium sulfate. During the month-long test oxidation levels in the system will be closely monitored for any observed difference in oxidation levels.

Materials Evaluation

Sample coupon racks containing several polymer- or rubber-coated specimens and various stainless steel coupons have been installed in numerous locations throughout the system. Additionally, pipe spool samples have been installed in the bleed stream and thickener underflow line. These spools are constructed of various steels with polymer or rubber linings.

Some of these corrosion samples will be removed after 6 months and the remainder at the end of 1 year. Recommendations for materials for future installations will be based on the analyses of these samples.

Sludge Disposal

A study of the effects of the long term disposal of the sludge generated by the dual alkali process has been developed. For the test program, unfixed sludge and two different combinations of sludge, fly ash, and quicklime will be placed with and without mechanical compaction in six separate impoundments (each 50 ft x 10 ft x 5 ft deep) for close study.

Test work will include leachate sampling, as well as a number of engineering and durability tests to characterize the sludge and sludge-mixture properties.

Centrifuge Evaluation

Because of the problems associated with operation of the filters in conjunction with a thickener, a pilot size centrifuge will be installed in early fall for experimentation. The centrifuge will be tested to determine its ability to produce an acceptable waste while separating sodium compounds from the cake. It will also be evaluated and compared with the rotary vacuum filters in terms of reliability and maintenance requirements.

CONCLUSION TO DATE

As indicated by operation since March 1980, and the successful completion of the acceptance test in July, the dual alkali process is capable of achieving greater than 90% SO_2 removal with an availability of more than 99% while processing a flue gas generated in a high-sulfur (>3.5%) coal-fired, full-size (280 MW) utility boiler. Consumption of raw materials and power was less than expected (guaranteed) while the SO_2 removal was over 94% on the average for the 12-day acceptance test.

Most of the problems initially encountered were mechanical and have been solved or greatly reduced in the operation at Louisville Gas & Electric's 280 MW Cane Run Unit 6.

Further investigation of filter operation, reactor operation, filter cloths, materials of construction, and major process component characterization is underway.

REFERENCES

1. Van Ness, R. P. et al., "Full-Scale Dual Alkali Demonstration System at Louisville Gas and Electric Co. - Final Design and System Cost," EPA-600/7-79-221b, NTIS No. PB80 146715, September 1979.

2. Van Ness, R. P. et al., "Project Manual for Full-Scale Dual Alkali Demonstration at Louisville Gas and Electric Co. - Preliminary Design and Cost Estimate," EPA-600/7-78-010, NTIS No. PB278722, January 1978.

3. Kaplan, N., "Summary of Dual Alkali Systems," in Proceedings: Symposium on Flue Gas Desulfurization Las Vegas, Nevada, March 1979, Volume II, EPA-600/7-79-167b, NTIS No. PB80-133176, July 1979.

OPERATING EXPERIENCE WITH THE FMC DOUBLE ALKALI PROCESS

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This paper reviews the design and initial operating experience with the flue gas desulfurization system at Southern Indiana Gas and Electric Company's (SIGECO's) A. B. Brown Station Unit #1, a 265 MW steam electric station burning up to 4.5% sulfur coal in a pressurized, pulverized coal boiler.

After initial checkout in the spring and summer of 1979, the FGD system began routine continuous operation. Overall operating results for sulfur dioxide collection, chemical consumption, availability, maintenance requirements, and operating costs are presented. The problem areas that contributed significantly to maintenance requirements or non-availability of the system are discussed in detail. Not counting the scheduled outage, the system has enjoyed a 96% availability overall in its first year of operation on a high sulfur coal application. Sulfur dioxide removal of over 90% has been routinely demonstrated. Overall operating costs on an annual revenue requirements basis are close to the original projections.

OPERATING EXPERIENCE WITH THE FMC DOUBLE ALKALI PROCESS

SYSTEM DESCRIPTION

The FGD system at Southern Indiana Gas and Electric Company's (SIGECO) A. B. Brown Station Unit #1 utilizes FMC's patented concentrated double alkali process for sulfur dioxide control. Figure 1 is a schematic representation of the process. The central purpose of this paper is a reliability analysis, for which we have chosen to divide the system into five major systems, some of which are overlapping:

A. Sulfur dioxide absorption
B. Lime chemical addition
C. Regeneration
D. Soda ash chemical addition
E. Sludge removal and disposal

Design Criteria

The A. B. Brown Unit #1 is a 265 MW steam electric station burning up to 4.5 percent sulfur coal in a pressurized, pulverized coal boiler. Make-up water to the FGD system comes from a collector well located adjacent to the Ohio River. Coal is transported to the site by rail car.

For equipment sizing and redundancy purposes the design basis is keyed to 23,788 m^3/minute (840,000 acfm) of flue gas at 138°C (280°F) for gas handling purposes and 85% collection of 9227 kg/hr (20,300 lb/hr) of sulfur dioxide for chemcial capacity.

Sulfur Dioxide Absorption

In the double alkali process, sulfur dioxide is absorbed according to the following reaction:

$$Na_2SO_3 + SO_2 + H_2O \text{ ---> } 2NaHSO_3$$

sodium sulfite + sulfur dioxide + water ---> sodium bisulfite

An important additional reaction is the oxidation of sodium sulfite:

$$Na_2SO_3 + 1/2\ O_2 \text{ ---> } Na_2SO_4$$

Sodium sulfite + oxygen ---> sodium sulfate

The sulfate ion, which is not active in absorbing sulfur dioxide, can be partially precipitated by reaction with calcium hydroxide. The remaining sodium sulfate is purged from the process through the entrainment of solution in the dewatered calcium sulfite/sulfate solids sent to the landfill.

The sulfur dioxide absorption is accomplished in the vendor's proprietary absorber. This absorber is designed to allow high sulfur dioxide collection efficiencies at a relatively low pressure drop without

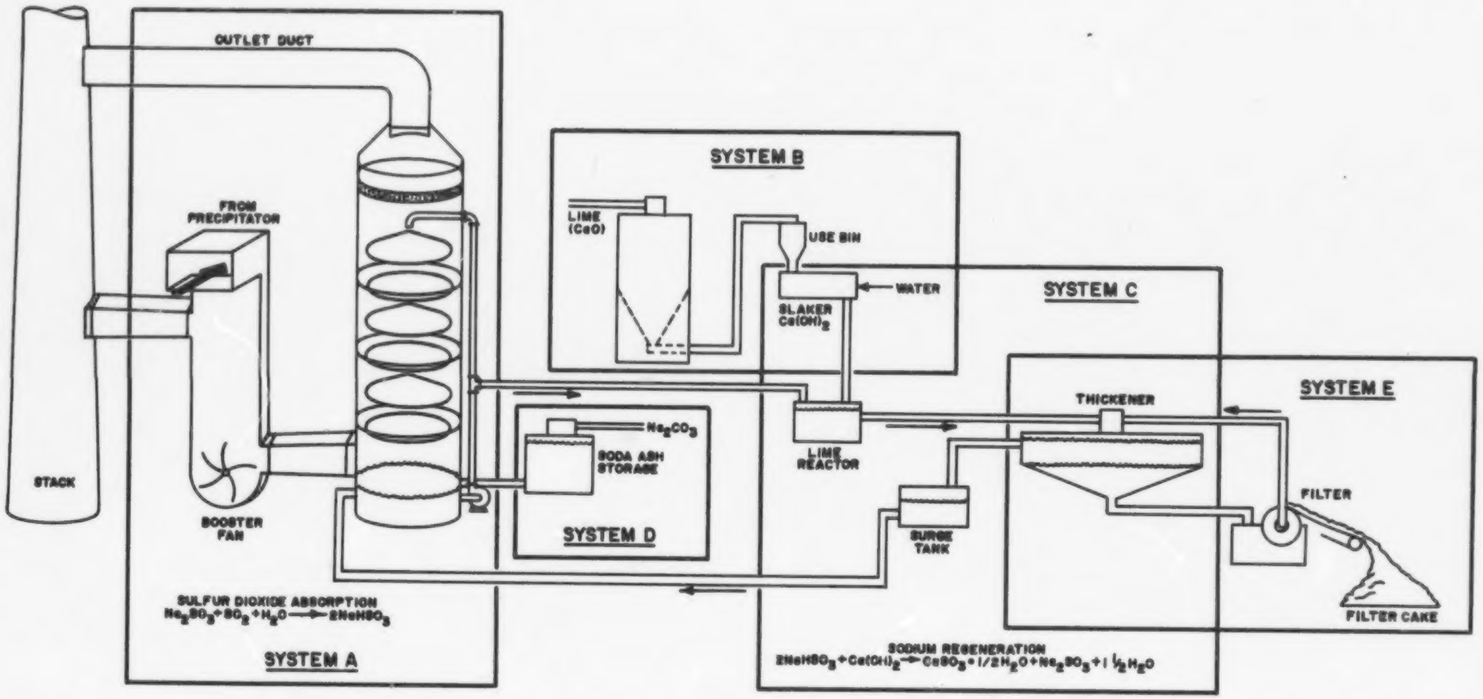

OUTLET DUCT
FROM PRECIPITATOR
STACK
BOOSTER FAN
SULFUR DIOXIDE ABSORPTION
$Na_2SO_3 + SO_2 + H_2O \longrightarrow 2NaHSO_3$
SYSTEM A
SYSTEM B
LIME (CaO)
USE BIN
WATER
SLAKER $Ca(OH)_2$
SYSTEM C
LIME REACTOR
Na_2CO_3
SODA ASH STORAGE
SYSTEM D
SURGE TANK
THICKENER
SYSTEM E
FILTER
FILTER CAKE
SODIUM REGENERATION
$2NaHSO_3 + Ca(OH)_2 \longrightarrow CaSO_3 \cdot 1/2H_2O + Na_2SO_3 + 1\,1/2\,H_2O$

Figure I. Double Alkali process schematic including duct arrangement.

the use of spray nozzles. Collection efficiencies above 92 percent have been demonstrated while operating at less than 12.7 cm (5 inches) of water pressure drop.

The pH of the scrubbing solution is controlled at 6.5. At this pH, the scrubbing solution contains equimolar concentrations of sodium sulfite and sodium bisulfite. This equimolar solution is highly buffered and can accept rapidly changing flue gas inlet conditions caused by swings in boiler load and/or changes in coal composition without upsetting the process control. As sulfur dioxide is absorbed, the ratio of bisulfite to sulfite increases causing a decrease in pH. A bleed stream from the absorber recirculation loop is directed to the lime reactor, and the absorber reservoir is replenished with regenerated sodium sulfite which maintains the scrubbing solution pH at 6.5. Maintaining pH levels in a range of 6.2 to 6.8 is important for several reasons. At a pH above 7.0, carbon dioxide absorption becomes significant and can lead to carbonate scaling. At a pH below 6.0, the vapor pressure of sulfur dioxide increases dramatically and can lead to equilibrium-inhibited sulfur dioxide collection.

Each absorber is about 9.14 m (30 feet) in diameter and 21.9 m (72 feet) tall to the outlet duct. Superficial gas velocity is approximately 2.7 m/sec (9 feet/sec) at design conditions. There are three stages of discs and doughnuts in each absorber. A schematic of the absorber internals can be seen in Figure 1. The bottom 2 m (7 feet) of the absorber comprise an integral reservoir for the recirculation liquor. To minimize wet/dry interface corrosion problems, the bottom disc and inlet plenum are made of Hastelloy G. The rest of the absorber internals are carbon steel lined with a glass flake polyester resin mastic. The absorber reservoir is additionally lined with acid resistant brick up to the bottom doughnut for thermal protection of the lining. A single stage thermoplastic chevron mist eliminator is provided downstream of the last absorption stage.

The recirculation liquor liquid-to-gas ratio at design condition is approximately 1.34 L/m^3 (10 gallon/1000 ACF). Recirculation liquor flow from the integral reservoir to the top of the last disc is provided by a rubber-lined centrifugal pump, which also provides bleed flow to the lime reactor.

A small slipstream from the recirculation liquor line is passed through a pH electrode to monitor and control recirculation liquor pH by controlling the regenerated liquor return flow. All of the recirculation liquor piping is fiberglass reinforced polyester for corrosion and abrasion resistance.

One of the interesting features of the system is the open bypass arrangement in which the ducting is designed to direct the gas to the system booster fans or through an undampered bypass duct directly to the stack. The advantage of this "open" bypass is threefold. First, it allows upsets in gas flow through the system to occur without affecting the boiler draft controls. Second, due to the high collection efficiency of the absorbers, it allows partial bypass of flue gas while maintaining compliance emissions; this minimizes chemical consumption while providing up to 11°C (20°F) reheat. Third, gas flow changes can be more readily accommodated because of the minimization of the number of dampers. Each absorber module is provided with guillotine isolation dampers in the inlet and outlet ducts. A louver damper is provided at the inlet to each booster fan for gas flow control. Each booster fan is capable of providing 14,273 m^3/minute (504,000 ACFM) at 17.8 cm (7.0 inches) of water pressure.

Lime Chemical Addition

Two pebble lime storage silos, each 9.14 m (30 feet) in diameter and 26.5 m (87 feet) tall and fabricated from carbon steel, provide 14 days supply at design conditions. Pebble lime is transported to the site by rail or truck. It is transferred from either storage silo via a pressure pneumatic system to any of three use bins, one above each slaker. Each use bin is fabricated from carbon steel and holds approximately one-half hour's supply at design use rates.

Regeneration

Calcium sulfite is precipitated by lime addition to regenerate sodium sulfite for use in the absorber according to the following reaction:

$$2NaHSO_3 + Ca(OH)_2 \dashrightarrow CaSO_3 \cdot 1/2\ H_2O + Na_2SO_3 + 1\ 1/2\ H_2O$$

Sodium bisulfite + calcium hydroxide --->
Calcium sulfite + sodium sulfite + water

The regeneration is accomplished in a low-residence-time continuously stirred tank reactor, which is controlled at a pH of 8.5, the titrametric endpoint of sodium bisulfite. The sensitivity of the pH control system is excellent at this set point resulting in effectively stoichiometric consumption of lime but a relatively wide control band.

The reactor is agitated with a vertically mounted top entry turbine agitator. The lime is fed to the reactor from two paste-type slakers each capable of feeding nearly 4990 kg (11,000 pounds) of $Ca(OH)_2$ per hour as approximately a 20 weight percent slurry. A third installed slaker provides a 100 percent spare for one of the other two slakers. Each slaker has an integral grit removal chamber. The reactor is provided with two immersion-type pH electrodes (one serves as an installed spare) which monitor and control the lime reactor overflow pH by controlling the feedrate to the slakers.

The lime reactor overflows to a 30.48 m (100 foot) diameter thickener tank where gravity settling of the calcium sulfite slurry takes place. The thickener concentrates the 1 to 2 weight percent solids in the feed slurry to 20 to 30 weight percent in the thickener underflow.

The regenerated liquor overflow from the thickener flows to the surge tank by gravity. Water is added to the tank by level control to maintain system water balance. Regenerated liquor is returned to the absorbers by a centrifugal pump. There is a 100 percent installed spare regenerated liquor return pump.

Soda Ash Chemical Addition

Soda ash is stored in a wet system. It is unloaded from a truck at a maximum rate of 9072 kg/hr (20,000 lbs/hr) into a proprietary apparatus for converting dry soda ash into the monohydrate crystal form in preparation for use as a saturated solution. Specifically, saturated soda ash solution from the tank is sprayed at about 1893 L/min (500 GPM) to wet the incoming dry soda ash. The wetted soda ash drops into the tank forming a bed of sodium carbonate monohydrate crystal in a saturated sodium carbonate solution. As soda ash solution is used in the process, fresh make-up water is added to the tank, dissolving the crystal bed to maintain a saturated solution. The advantage of using a saturated solution for chemical make-up is that it allows sodium addition to the absorbers to be controlled by the volume of saturated liquor delivered to the absorbers because the concentration is constant. Soda ash solution from the soda ash storage tank to the absorbers flows continuously in a loop to minimize concentration and temperature gradients within the solution layer above the crystal bed in the storage tank. This also helps prevent crystallization in the soda ash transfer lines. In addition, all soda ash piping is heat traced and the tank is also provided with steam plate coils in order to maintain solution temperature. Flow through the transfer loop and to the absorbers is provided by one of two centrifugal pumps.

Sludge Dewatering and Disposal

The sludge dewatering equipment consists of three rotary vacuum filters, each sized for 33 1/3 percent of total capacity required when burning the maximum sulfur coal (4.5 percent). When burning the nominal coal (3.7 percent sulfur) each filter is essentially a 50 percent filter. Thickener underflow is pumped to the filter vats by an air operated diaphragm pump. There are two full flow underflow pumps installed per filter. The rotary vacuum filters are primarily of carbon steel construction. The filters are knife-discharge type, and the 50 to 60 percent solids cake is discharged directly into dump trucks for transportation to the on-site landfill area. Each filter is equipped with a wash belt compression assembly for applying wash water to the cake to enhance sodium recovery.

SYSTEM AVAILABILITY

Overall system availability as defined by PEDCO for the first 13 months of routine operation beginning in August, 1979, is summarized in Table 1. While we feel that the definitions of some of the PEDCO parameters leave something to be desired from a utility point of view, they at least provide a consistent basis for comparison.

Table 2 shows the incidents that contributed to system unavailability for the same period. The total scrubber forced outage rate was 3.3%. At SIGECO, a forced outage rate for the boiler and turbine of 1% is considered good, and the goal at A. B. Brown. There exists here a good comparison of boiler-turbine to scrubber state-of-the-art design. Forced outages on the scrubber occur at a frequency of three plus times what we strive for on the remainder of the unit. In addition to the normally scheduled annual outage, we feel that an additional outage is required each year for inspection of ducts, linings and breechings, due to the possibility of corrosion.

TABLE 1 PEDCO INDICES

	Availability	Operability	Reliability	Utilization
August, 1979	96	93	93	93
September	99	76	88	76
October	88	99	99	86
November	97	96	97	88
December	81	81	81	81
January, 1980	98	98	98	98
February	99	70	98	69
March	45	65	68	27
April	100	100	100	96
May	100	100	100	83
June	100	100	100	98
July	92	93	93	88
August, 1980	95	95	95	95

TABLE 2 SOURCES OF UNAVAILBILITY

August 1, 1979 to August 31, 1980

Unavailability Cause	Hours
Recirculating Pump Failure (2)*	70
Miscellaneous Electrical Trips (4)*	6
Slaker Feeder Controls	12
Thickener Rake Stall or Overload (3)*	149
Isolation Dampers	3
Lime Reactor Overflow Elbow	10
Thickener Underflow Pump Suction Plugged (2)*	37
Lime Transfer System Plug-up	3
Rotary Filters Unavailable	18
Low Water Pressure	2
Subtotal	310
Scheduled Outages	594
Total	904

Period Hours	9528
Availability	90.5%
Discounting Scheduled Outages	96.7%

*Denotes number of incidents, if greater than one.

The following section provides a more detailed chronological discussion of the incidents that contributed to unavailability.

Month-By-Month Performances

The period of March through July, 1979, constituted the initial check-out and debugging of the system. There were some areas which required modifications for mechanical improvements to cope with situations not envisioned during the original design phase. Details on these modifications were presented in an earlier paper by Durkin et al (1).

The system availability began to improve dramatically in August as operator attention and awareness increased and as mechanical problems became fewer. The system was down the first day of August for a recirculation pump impeller lining failure. The only other system outage was for the removal of tramp metal from the lime transfer system. These two items account for 32 hours of down-time.

September's record shows 48 hours down for an FGD system water filter tie-in, necessitated by high suspended solids in the service water. A failure of a 460 volt power cable to one of the recirculation liquor pumps caused a 600 amp main feeder to overload and trip. It took only two hours to reset the feeder and restart all of the systems. The spare recirculation liquor pump was put in service and full FGD system operations resumed. For two separate periods totalling 51 hours, the FGD system was bypassed so that precipitator particulate emissions could be tested. There was one short booster fan trip caused by a mis-wired relay which was not to be diagnosed until several weeks later. At the end of the month, the filter cake quality got so poor that the dump trucks could not handle it. We elected to shut down the FGD system so that the filter building could be cleared of spilled cake.

The filter building clean-up continued into the first day of October. Again, a short duration booster fan trip occurred. On October 28, the FGD system and boiler were taken off-line for a scheduled precipitator and FGD system inspection.

The inspection outage continued through the first three days of November. Once put on-line, the only system outage was due to electrical problems of the slaker control printed circuit boards. It took about 12 hours for the cause to be diagnosed and repairs made.

The most significant outage occurred in December. After a short duration boiler trip, the thickener rake was discovered to be stopped. Numerous attempts were made to restart the rake to no avail. Finally, the thickener had to be drained and the solids removed to free the rake. The system outage lasted 146 hours.

January's record shows only 12 hours of downtime. The thickener rake torque instruments indicated a steady increase almost to the point of motor overload. Since we were fearful of another stall, the system was shut down. After several hours of filtering, the torque dropped and the system put back on line.

We experienced a period of poor quality filter cake in February, similar to September's situation. We elected to take the FGD system off-line for building and roadway clean-up.

March downtime was split between two outages on the balance of plant and the completion of roadway work begun in February. Availability was 45% based on 444 boiler hours.

May downtime was due to an outage on the make-up water line to the cooling tower during which some internal scrubber inspection work was done.

July downtime was caused by two scrubber recirculating pump impeller liner failures in three days.

August downtime was caused by pluggage of the thickener underflow pump suction header, and a couple of other minor problems.

Overall, since the FGD system reached stable operation early in August of 1979, it has operated at 90% availability and a 3.3% forced outage rate. The FGD system's longest continuous run has been 54 days, and it had run 71 days between partial forced outages.

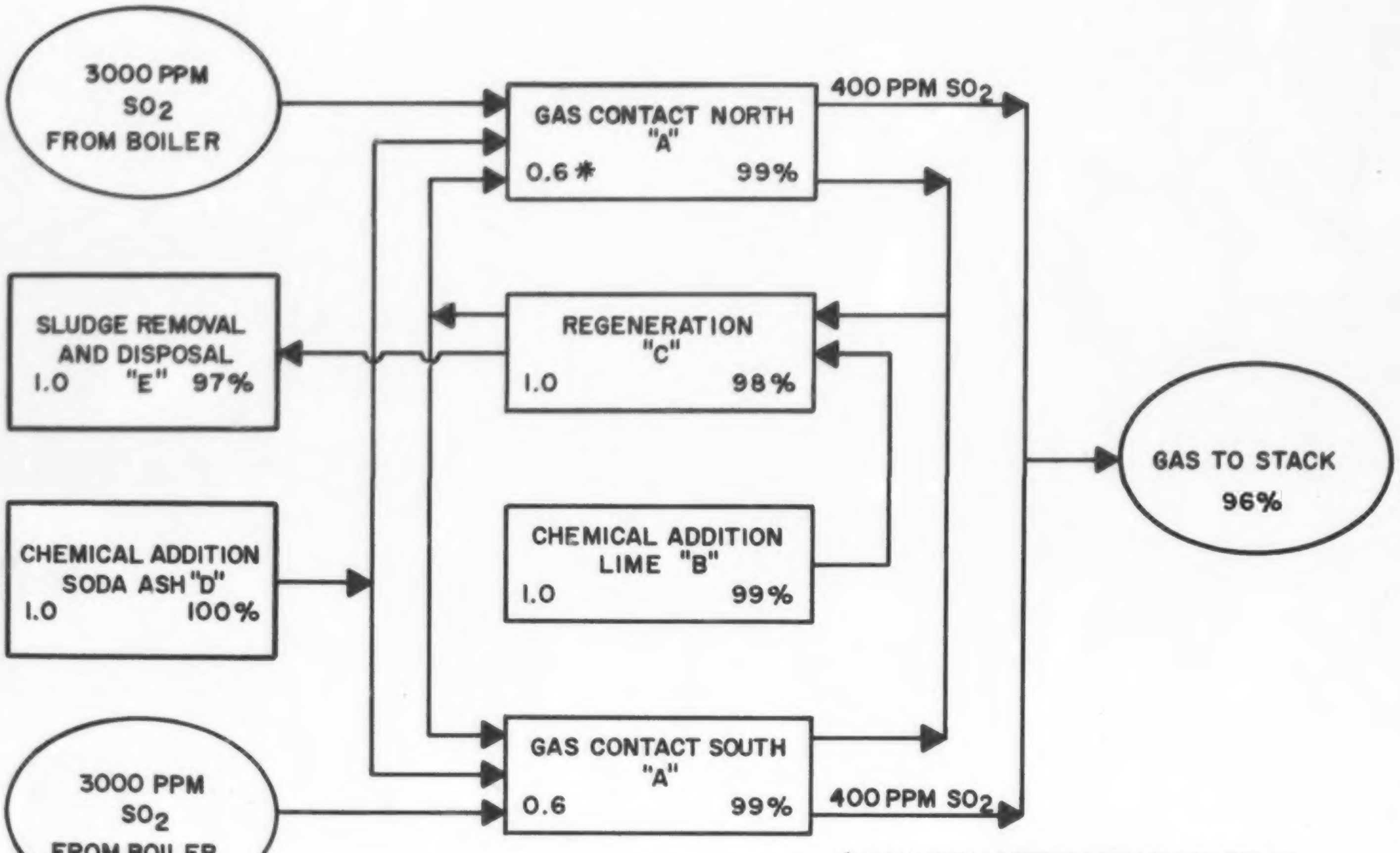
RELIABILITY ANALYSIS

While the total availability record at A. B. Brown has been good compared to most other FGD systems now in operation, our goal is to achieve a forced outage rate of 1%. To better understand how we can improve our availability in this and future installations, we have performed a reliability analysis on each of the five major systems described in the first section of this paper. The systems are not mutually exclusive, meaning that an element may be included in more than one system. The thickener tank is an example, since it is included in both the regeneration system and the sludge removal system. Also, the systems do not necessarily follow the flow of one fluid, but generally follow the series of events or reactions that must occur to insure availability and compliance.

Figure 2 shows the block diagram of the five systems. All of these systems are integral to scrubber availability and the failure of any system will result in scrubber downtime. Reduced capacity of a system may result in unavailability, but most likely will be considered available at low load. With a total FGD system availability of 96%, shown at the far right side of Figure 2 the contribution of each system is shown in the lower right hand corner of each box. The decimal in the lower left hand corner is the sizing of each part, compared to the boiler full load gas flow or maximum sulfur dioxide collection.

Sulfur Dioxide Absorption (A)

Figure 3 illustrates the elements of the sulfur dioxide absorption system. The gas contact systems have performed at a reasonable level of reliability, 99% plus, but not without some problems. The absorbers have operated perfectly, with no pluggage and only minimal lining deterioration. The single stage chevron type mist eliminator has given us some problems. In


3000 PPM
SO_2
FROM BOILER
SLUDGE REMOVAL AND DISPOSAL
1.0 "E" 97%
CHEMICAL ADDITION SODA ASH "D"
1.0 100%
3000 PPM
SO_2
FROM BOILER
GAS CONTACT NORTH "A"
0.6 * 99%
REGENERATION "C"
1.0 98%
CHEMICAL ADDITION LIME "B"
1.0 99%
GAS CONTACT SOUTH "A"
0.6 99%
400 PPM SO_2
400 PPM SO_2
GAS TO STACK
96%

* DENOTES SUBSYSTEM FRACTION OF TOTAL SYSTEM DESIGN. CAPACITY.

Figure 2. Reliability Block Diagram of FGD major systems.

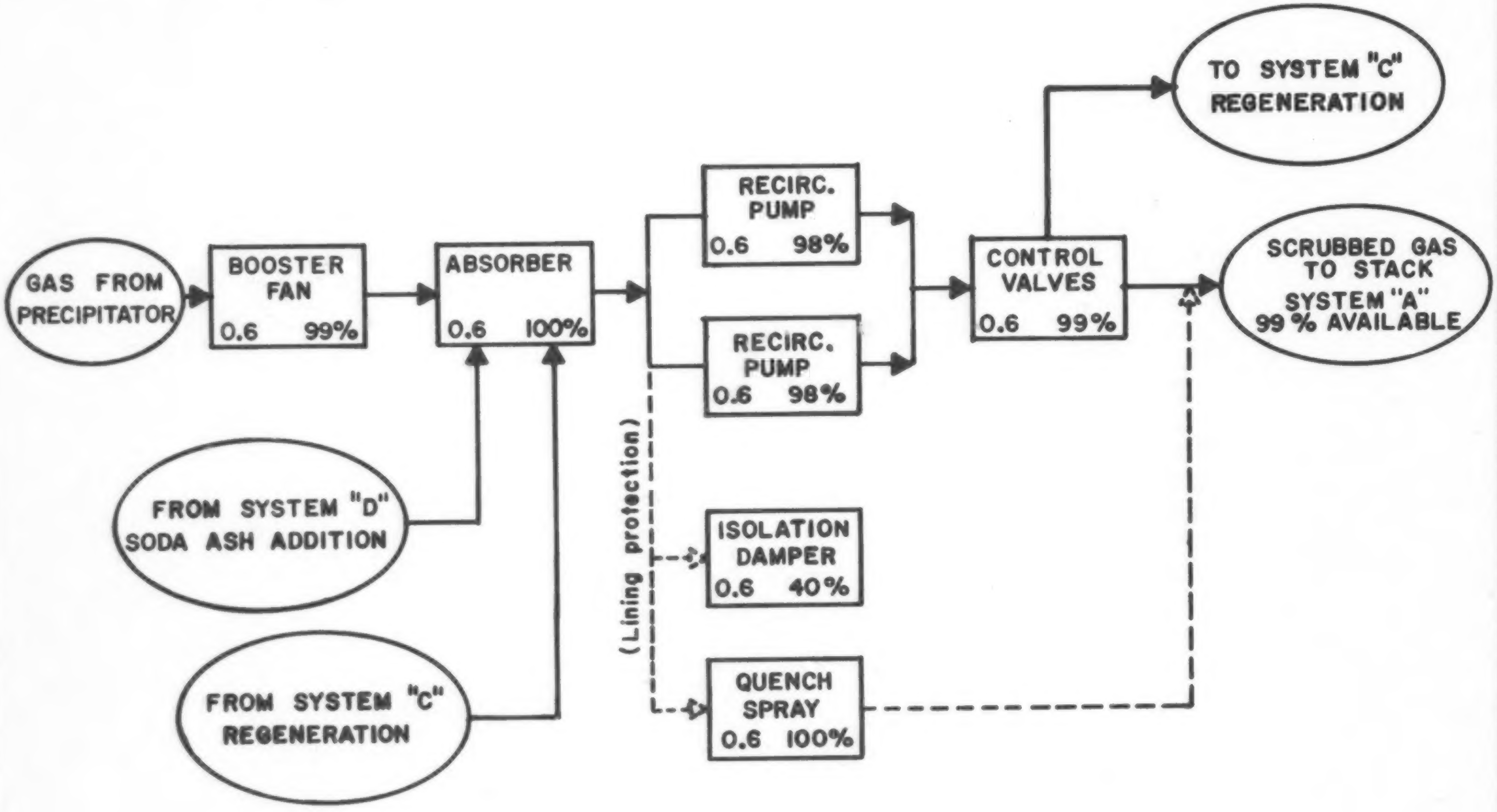

GAS FROM PRECIPITATOR
BOOSTER FAN 0.6 99%
ABSORBER 0.6 100%
FROM SYSTEM "D" SODA ASH ADDITION
FROM SYSTEM "C" REGENERATION
RECIRC. PUMP 0.6 98%
RECIRC. PUMP 0.6 98%
(Lining protection)
ISOLATION DAMPER 0.6 40%
QUENCH SPRAY 0.6 100%
CONTROL VALVES 0.6 99%
TO SYSTEM "C" REGENERATION
SCRUBBED GAS TO STACK SYSTEM "A" 99 % AVAILABLE

Figure 3. System "A", gas contact (one side only)

June of this year, recirculating liquor pumps in the south module began to fail at an alarming rate. Cavitation was the suspected problem, caused by pluggage of the pump suction. Inspection revealed that two sections of the mist eliminator had collapsed into the scrubber sump, blocking the suction pipe. The mist eliminator sections appeared to slip off the 7.5 cm (3 in) wide shelf supporting them, and the Inconel tie-wires were not adequate to hold them.

The mist eliminators seem to perform adequately when in place. However, in recent months there has been evidence of increasing losses of sodium due to entrainment. The problem appears to coincide with the collapse of several sections of the mist eliminator. The losses are primarily in the form of high sodium levels in the drainage from the stack. Sodium levels in the exit gas, as determined by analysis of EPA Method 5 particulate catches, have been consistently low, in the range of 0.005 to 0.006 lb/mm BTU, or about 10% to 15% of the total particulate. This suggests that the entrained drops are large enough that they fall down the stack. Stack drain losses of sodium may amount to 1 to 2% of the SO_2 collected, or up to more than one-half of the total excess sodium consumption, which will be discussed in more detail in a later section. Steps are now being taken to repair the existing mist eliminator and add a second level of eliminators to reduce losses.

The rubber-lined recirculating liquor pumps have performed about as expected. During engineering, pump lining and impeller life were estimated at one year and the pumps were spared accordingly. Disregarding cavitation problems and some sub-par work on original installation, our experience has been reasonable.

The system design calls for three levels of protection for the scrubber lining: 1) cooling with recirculated liquor with the recirculation pumps; 2) isolation dampers; 3) water quenching of inlet gas. Plant experience with the guillotine isolation dampers has been poor. Original materials of construction caused some problems, and of late, seemingly minor problems such as a stuck relay in the entry door logic have led to major problems. In addition, the lead time for fabricated pieces of a material suitable for a wet flue gas environment is truly ridiculous. As a result, the plant relies on water quenching of the incoming gas to protect scrubber internals more than we rely on dampers.

The outlet duct from the absorber to the outlet isolation damper is coated with a flake glass lining. In the outlet duct from the damper to the stack, which is lined with an epoxy vinyl ester coating, the lining has failed dramatically on one module, subjecting the carbon steel structure to severe corrosion. This duct segment is exposed to cool 54°C (130°F) saturated gas when the scrubber is in service and to hot 149°C (300°F) gas those few times it is off. An appropriate low carbon, high molybdenum stainless steel will be installed at the next scheduled outage as a replacement of the corroded duct. The corresponding duct segment on the other module will be lined with a new experimental lining we want to try. This duct section was patched from the outside with the unit on line, and has caused no downtime.

Lime Chemical Addition (B)

The lime chemical addition system shown in Figure 4 has been responsible for less than 1% of the FGD system 3.3% forced outage rate, but this 99% availability is not an accurate measure of the strength of the design and operating capabilities of this system. There have been numerous instances of reduced capacity because of inadequate lime supply to regeneration. However, the surge capacity of the thickener tank has helped us minimize any lost time or non-compliance. The system can run for several hours before a reduction in scrubbing capacity, due to the storage of 3,028 kiloliters (800,000 gallons) of regenerated liquor in the thickener. Circulating low pH (6.5) liquor through unlined carbon steel piping and tanks does cause some accelerated corrosion and at every outage great care is taken to ascertain material integrity with a view toward possible future replacement. To date, no problems have been seen, but we are installing additional monitors to record the pH in these unprotected areas. As our experience grows, we will be establishing some lower range cutoff points below which we will not operate, based on engineering judgments.

Interruptions in lime supply are caused by failure of the lime transfer system or by foreign material in the lime. The transfer system itself, while it has performed reasonably well, is recognized by plant personnel as a weak link. The system has only one blower and one feed line going from the two storage silos to the slaker use bins. A malfunction of any one component will shut down the entire system, and result in low pH incidents described previously. We are engineering some redundancy for lime transfer to the slakers. As a system, lime transfer has been only 98.5% available.

Regeneration (C)

The second largest cause of scrubber forced outages is the regeneration system. FIgure 5 shows the elements of the system. Slakers are included in both system B, lime addition, and C, regeneration. The regeneration area has not caused much downtime on the scrubber, but as with the lime addition system that is not a fair measure of how well it performed. Our incidence of reduced capacity due to regeneration problems has been significant.

As designed, pH at the lime reactor discharge is fed back to the slaker feeder controls to control lime addition. The pH normally swings plus or minus one unit from the setpoint. The control problems in this situation are obvious. As a result, slaker feeders are constantly varying in response to both load changes and normal swings. This causes greater than normal wear on the rotary air lock feeders, and they overcharge the slakers occasionally, causing plug-ups. Our early experience with lime feeder controls was very poor. Slaker control problems are often attributable to the dusty, wet environment of the slaker building. All local control panels are mounted adjacent to the slakers, and subjected to the same ravages of steam, caustic, and water as the slaker itself. NEMA dustproof rated enclosures have brought these problems to a manageable level.

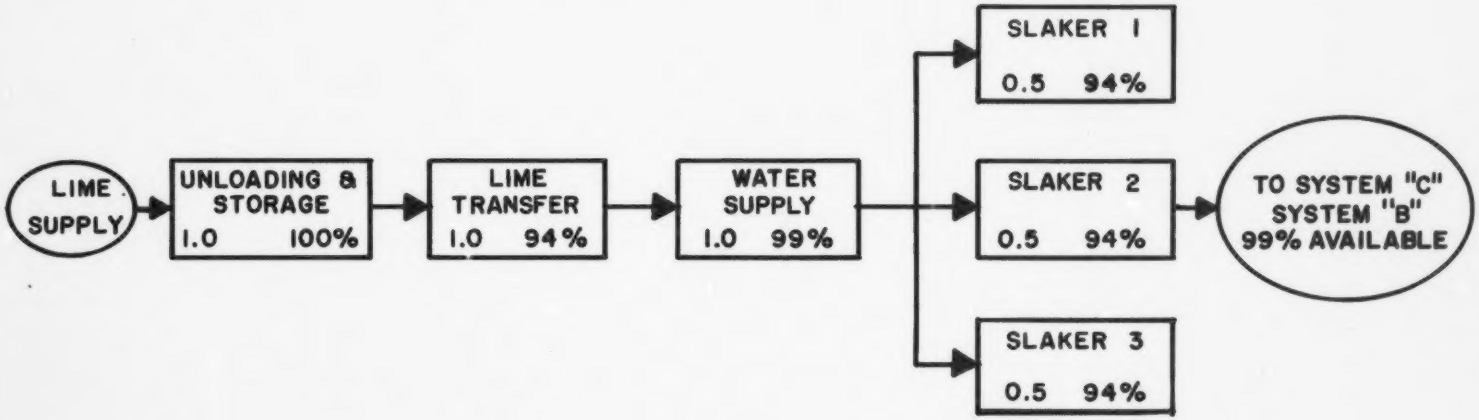

LIME SUPPLY
UNLOADING & STORAGE
1.0 100%
LIME TRANSFER
1.0 94%
WATER SUPPLY
1.0 99%
SLAKER 1
0.5 94%
SLAKER 2
0.5 94%
SLAKER 3
0.5 94%
TO SYSTEM "C"
SYSTEM "B"
99% AVAILABLE

Figure 4. System "B", lime chemical addition

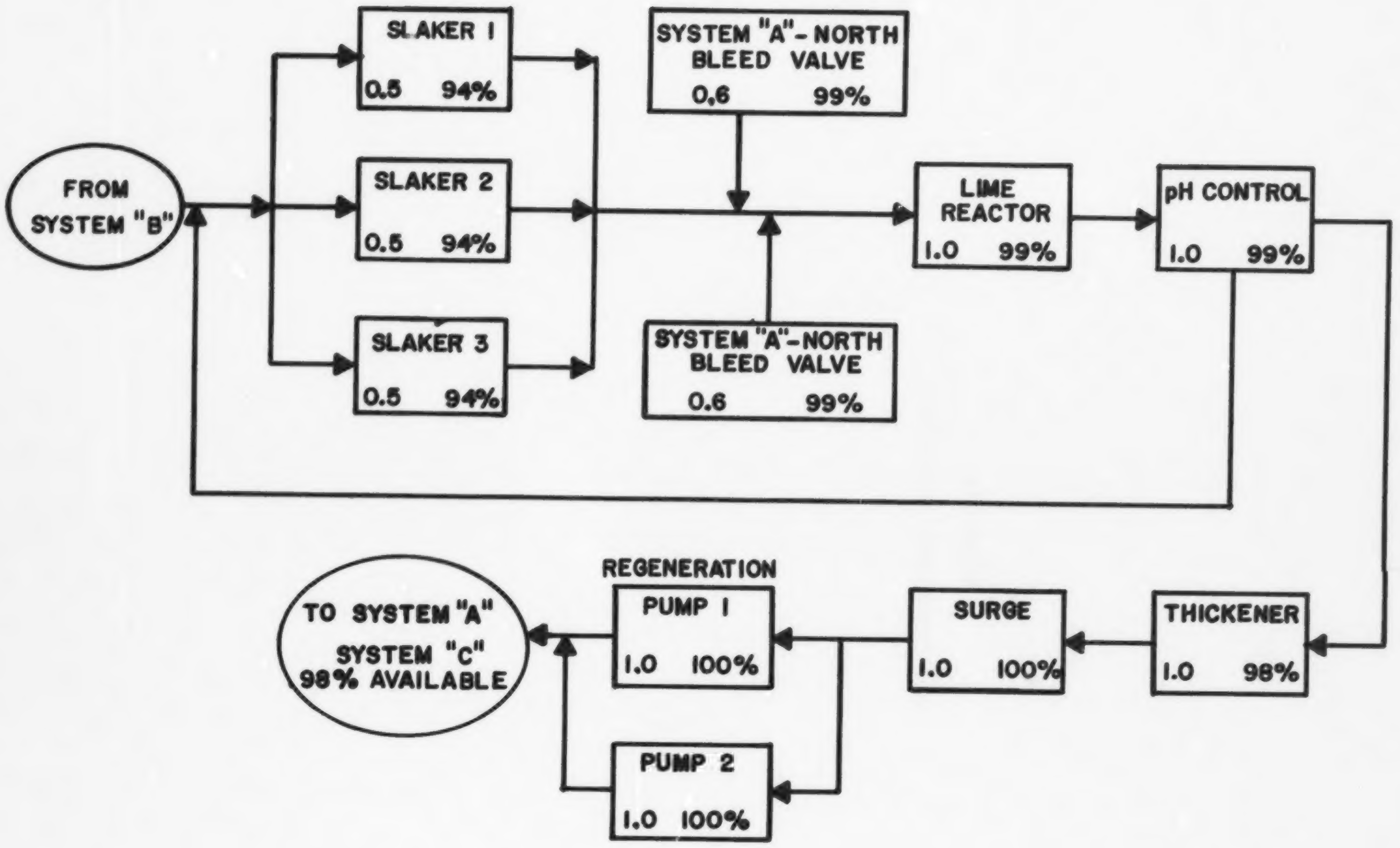

FROM SYSTEM "B"
SLAKER 1 0.5 94%
SLAKER 2 0.5 94%
SLAKER 3 0.5 94%
SYSTEM "A"- NORTH BLEED VALVE 0.6 99%
SYSTEM "A"-NORTH BLEED VALVE 0.6 99%
LIME REACTOR 1.0 99%
pH CONTROL 1.0 99%
THICKENER 1.0 98%
SURGE 1.0 100%
REGENERATION
PUMP 1 1.0 100%
PUMP 2 1.0 100%
TO SYSTEM "A" SYSTEM "C" 98% AVAILABLE

Figure 5. System "C", regeneration

The lime reactor overflow elbow has been patched several times and is scheduled for replacement in the near future. Failure analysis questioned the suitability of carbon steel in a service where it is exposed to abrasion from lime grit and corrosion from occasional pH excurisons.

The remainder of the regeneration system has operated well with minimal problems, and has offered the level of reliability that is expected from power plant machinery.

Soda Ash Chemical Addition (D)

The soda ash chemical addition system shown in Figure 6 has not contributed to any FGD system downtime, although low sodium concentration in the system has resulted in some non-compliance. This is not seen as a problem though, and with fourteen days' inventory available in the tank, no changes are foreseen.

Sludge Removal and Disposal (E)

As shown in Figure 2, the sludge removal and disposal system has been the weakest of the five systems, due to both mechanical and process difficulties. This weakness would be even more costly in terms of availability if the system did not have some surge capacity in the thickener tank, which allows the plant to run as long as eight hours at full load without filtering. However, this is not a preferred operating mode, and it is not without some detrimental system effects, so in deciding to utilize the thickener surge capacity the value versus the consequences of continued operation must be carefully considered.

The detrimental side effects of utilizing the thickener surge capacity are as follows. First, as part of our filter cake quality testing, it was shown that increasing the inventory of solids in the thickener tanks tends to increase the pH of the underflow, apparently due to the continuing reaction of small amounts of alkali. This increase in underflow slurry pH coincides with a deterioration in cake quality. Second, the thickener is designed with a pivoted rake and no powered rake lifting mechanism. The rake and drive are protected from overload by a spring loaded clutch, designed to trip the rake drive when it reaches its torque rating. If the rake were to trip with a high inventory of solids in the tank, it would sink into the mud and be impossible to restart. This would require a lengthy outage to pump out the thickener tank with the loss of $50,000 to $100,000 worth of chemicals. Third, operation of the gas contact system without filtering will increase the total volume of material -- solids and liquids -- in the thickener tank and result in liquid losses because of overflowing the surge tank capacity. This costs money in chemical losses, amounting to 10 to 20% of the excess sodium consumption.

Figure 7 shows in greater detail the reliability analysis of the sludge removal and disposal system, and shows elemental availabilities much lower than the 97% for the system. Redundancy is the key, along with high maintenance requirements and quick response to problems. Availabilities of the filters have been in the mid eighties, but the problems have been generally unrelated to design. Rotary vacuum filters are high maintenance

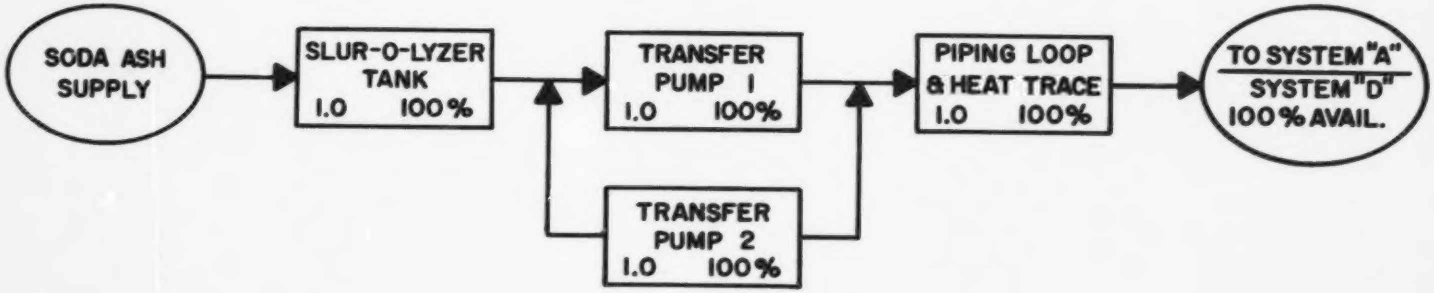
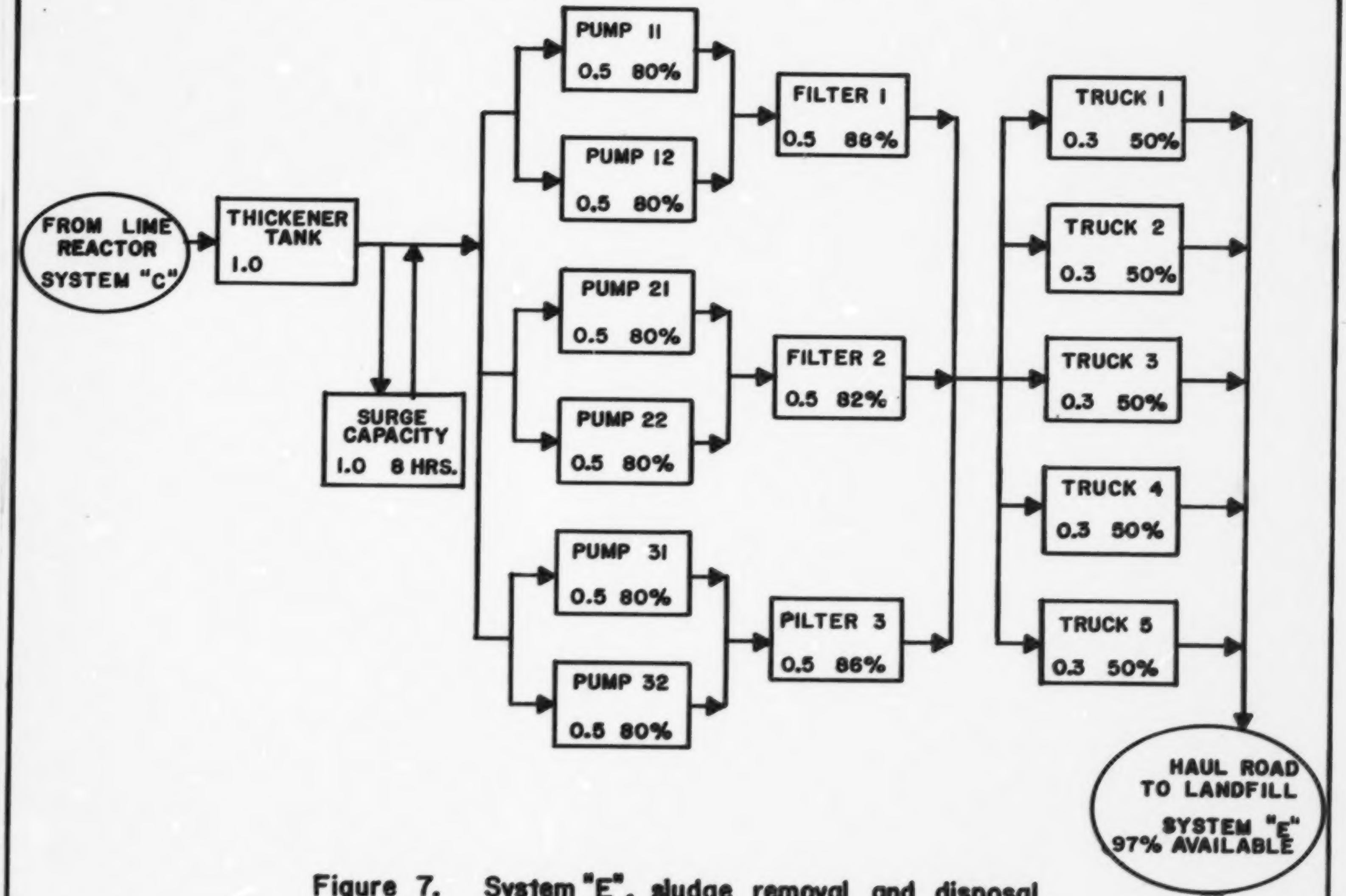

SODA ASH SUPPLY
SLUR-O-LYZER TANK 1.0 100%
TRANSFER PUMP 1 1.0 100%
TRANSFER PUMP 2 1.0 100%
PIPING LOOP & HEAT TRACE 1.0 100%
TO SYSTEM "A"
SYSTEM "D" 100% AVAIL.

Figure 6. System "D", Soda ash chemical addition.



FROM LIME REACTOR SYSTEM "C"
THICKENER TANK 1.0
SURGE CAPACITY 1.0 8 HRS.
PUMP 11 0.5 80%
PUMP 12 0.5 80%
PUMP 21 0.5 80%
PUMP 22 0.5 80%
PUMP 31 0.5 80%
PUMP 32 0.5 80%
FILTER 1 0.5 88%
FILTER 2 0.5 82%
PILTER 3 0.5 86%
TRUCK 1 0.3 50%
TRUCK 2 0.3 50%
TRUCK 3 0.3 50%
TRUCK 4 0.3 50%
TRUCK 5 0.3 50%
HAUL ROAD TO LANDFILL
SYSTEM "E" 97% AVAILABLE

Figure 7. System "E", sludge removal and disposal.

items. Cloth life is only one to three months, and between cloth changes, maintenance is required regularly to keep caulking ropes in place and to repair holes in the cloth. A large portion of our filter problems concern the vacuum pumps and filtrate return pumps becoming overloaded with solids carried over from the filter, generally through holes in the cloth. The solids appear to be mainly grit discharged from the slakers. The lime supply at 88% - 90% available CaO, has a reasonably high amount of grit which, no doubt, contributes to this condition.

Another problem area is the underflow pumps, which have had only 80% availability. Here the problems are basically design related. The pumps are air operated double diaphragm type. Air supply to the pumps is controlled by the filter vat liquid level. As vat level drops, the pumps are energized to refill vat level. This constant on-off operation allows the pump, the suction lines, and the discharge lines to lie full of thickener underflow slurry. They will eventually clog. Various operating procedures to flush the system on startup and shutdown have lessened the problem but not eliminated it. It appears that any design with low or intermittent flow in this critical area is a weak one.

The final element in the sludge removal and disposal system is the trucks. A. B. Brown is using tandem axle dump trucks to transport filter cake to the landfill. Two to four trucks are required for full load operation depending on the combination of filters in service. Our experience has been very poor in this area. With five trucks assigned to filter cake hauling, we have had three or more available for load only 75% of the time. This is another area where the surge capacity of the system comes into effect. It gives us enough time to repair either the right combination of filters or trucks to maintain availability and compliance.

Truck problems have been in two general areas: drive train (transmission and axles) and tailgate. Both of these are contributed to very heavily by the condition of the sludge. Wet, soupy material is not only hard to handle in the landfill but exceptionally hard to drive through, and it exerts a great deal of hydraulic pressure on the tailgates.

Trucks frequently must be taken out of service to clean beds. This situation occurs not just in winter when cold weather causes some freezing to beds, but also during hot weather. We have experimented with plastic liners and feel that they are an improvement but not a cure-all. Many of our truck problems could probably be solved with more suitable trucks. However, the existing filter building layout limits our choices.

In addition to the mechanical aspects of the sludge removal system performance, there are process considerations. There have been extreme variations in cake quality that are not entirely due to mechanical conditions. While difficult to quantify, it is clear that poor cake quality increases the cost of operation and maintenance and negatively impacts availability.

A large effort in the last six months has been devoted to understanding the variations in quality, and while we're not completely satisfied that we understand what's happening, we have found some general truths. First, it is clear that cake quality is directly related to crystal size and shape; moisture content is not so important. Second, as noted above, elevated pH generally deteriorates cake quality, at least in this application. Third, transient process conditions are bad; our worst cake has always been associated with situations in which we changed solution chemical composition rapidly.

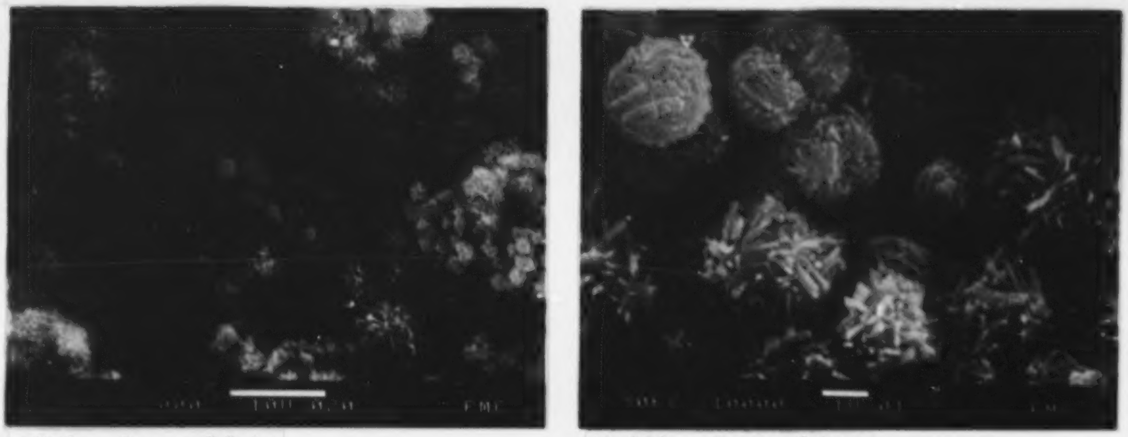
Finally, we believe there is a relationship between sodium sulfate concentration and cake quality. Sodium sulfate concentration seems to affect the size and shape of the calcium sulfite crystals formed in the regeneration section. All other things being equal, the greater the sulfate concentration the larger the size and the more irregular the surface of the crystal, as shown in Figure 8. Crystals with an irregular porcupine-like surface are called radial crystals. The larger, more radial crystals result in a filter cake with better handling properties. As noted earlier, sodium sulfate formed in the system is purged from the process through the entrainment of solution in the dewatered filter cake. Since the fuel sulfur level is high, there is a large amount of cake formed in this system. This means that the concentration of sodium sulfate in the scrubbing solution does not need to be very high to maintain equilibrium in the system. (The amount of sulfate formed is relatively fixed for a given load. Thus, the more cake formed the lower the concentration of sulfate in the cake.) The situation is aggravated by transient losses of sodium sulfate through carryover into the gas stream or spills.

There are three general categories of sodium loss to consider: entrainment, filter cake, and miscellaneous spills and leaks. Entrainment, or stack losses, was discussed earlier, and we hope to correct that problem during our next outage. Filter cake washing, while steadily improving, is still not up to design. We plan to increase our hot water capacity in the FGD area in an attempt to improve our washing. Spills, a problem in early operation, have been brought under control in recent months, but we still are considering increasing our surge tank capacity to allow for greater fluctuations in the water balance. We are hopeful, but still not certain, that these improvements will allow us to simultaneously reduce sodium consumption and improve cake quality.

While we endeavor increasingly to implement the above improvements in our routine operation, there are aspects of cake quality that we still do not understand to our satisfactaion. Continuing research projects at FMC's Central Engineering Laboratories and Purdue University, together with experiments at A. B. Brown, will hopefully eventually lead to a scheme for consistent production of easily handleable cake and a better fundamental understanding of the numerous parameters that effect the cake quality.

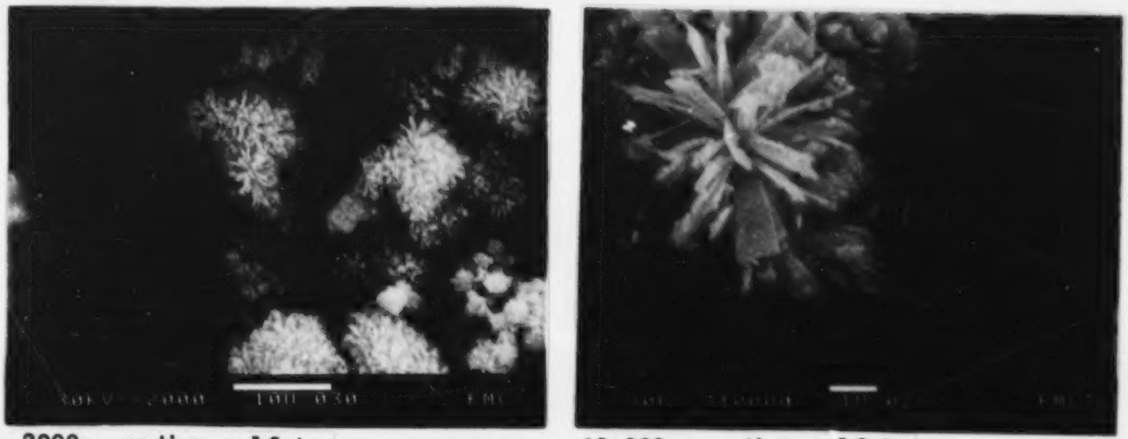
Conclusion

Typically, in the operating life of mechanical equipment, there is a break-in time during which breakdowns are frequent; a useful operating life during which breakdowns are at a low, manageable level; and finally a wear-out period when failures increase dramatically. Without changes



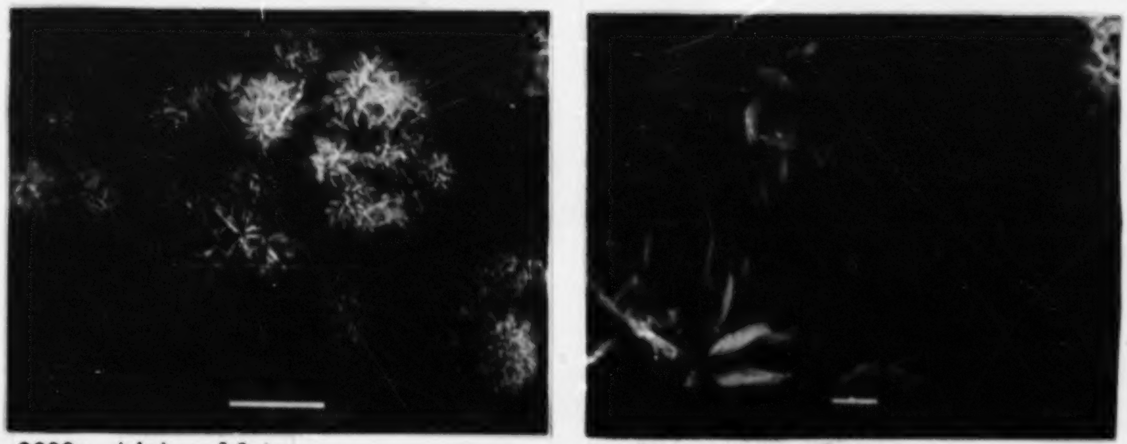
2000x, low sulfate

10,000x, low sulfate



2000x, medium sulfate

10,000x, medium sulfate



2000x, high sulfate

10,000x, high sulfate

Figure 8. Effect of increasing sulfate concentration on crystal structure.

and improvements, we feel that the availability of the equipment has reached its high point. With that in mind and remembering that our scrubber forced outage rate is three plus times what we would like, we are engineering system improvements in these areas:

1. Filter cake quality
2. Underflow pump recirculation
3. Duplicate lime transfer system
4. pH controls

We hope that these improvements will affect the wear-out period and allow us to improve our availability. Some other problem areas addressed in this paper are locked into the system because of original design, layout, or available space. Frankly, we feel that it is unrealistic to expect the first utility installation of a new FGD technology, even with the process advantages of double alkali, could ever achieve the level of reliability and availability that the balance of the system achieves. However, we feel that the problems we have had are primarily mechanical, and correctable, in future installations.

OPERATING COSTS

The variables in the FGD system operating costs are: operating manpower, maintenance expenses (labor and materials), lime usage, soda ash usage, power and landfill costs. In analyzing these costs for equation back to the manufacturer's original estimate, upon which the selection of the double alkali scrubber over a lime or limestone system was based, several adjustment factors must be applied to the costs incurred in order to put them on the same basis as the specification and proposal.

The system is designed to handle 265 MW gross of flue gas at 143°C (290°F) and 5.0% O_2. Coal burned was to be a maximum of 4.5% sulfur, 26,749,000 joules per kilogram (11500 Btu per pound) and 0.05% chlorides. Cost analysis was based on 70% load factor. Actual experience has been off-evaluation on some key items: sulfur has averaged much closer to 3.5%; the unit load factor has been 58% rather than 70%; and the flue gas volume and excess air have been higher than anticipated. All these items significantly impact the cost comparison of actual to guarantee. Rather than go through a laborious explanation of each variable with all applicable adjustments, it is most likely more informative to qualify each.

Operating manpower was based on one additional man per shift. Our experience has been very favorable in this area. In all but extremely unusual circumstances, one local FGD system operator has been sufficient to operate all equipment. The scrubber control board is located in the main control room, adjacent to the boiler-turbine-generator control panel, and this design feature has been a big manpower saver. Existing control room personnel operate the panel.

Maintenance costs (including electrical and instrument) were predicted to be 1.5% of capital costs per year. Both labor and material were included in the 1.5% figure. Our experience has shown this to be low by about 50% after allowing for inflation. Future maintenance costs are expected to increase as we approach the "wear out" period on some of our major pieces of equipment. The 1.5% maintenance was a concession made during bid analysis to the claimed features of the double alkali process versus 2.5% of capital for lime systems and 4.0% of capital for limestone systems. While our experience has not been quite that good, experience by others would indicate that our predictions for lime and limestone were also low.

Excess lime usage was predicted to be 1% or less (stoichiometric ratio ≤1.01). Our experience has shown excess lime usage as low as 0.02% (stoichiometric ratio = 1.002) to be attainable under normal operating conditions.

Soda ash usage was predicted to be 2.5% of the moles of SO_2 collected plus 1% allowance for the chlorides in the coal. As already discussed, our usage has been higher than expected and the increase over design has been attributed by FMC to inability to wash the filter cake with adequate amounts of hot water and to system losses through spills and stack drainage. It is anticipated that improvements in soda ash usage will be made. Soda ash prices have jumped sharply in the past year, and the price now is almost double what was budgeted in 1976. However, soda ash is not a large percentage of total operating cost and we are hopeful that recent price increases, which were caused by some unexpected closings of obsolete synthetic soda ash plants, will not recur.

Power requirements were predicted to be about 0.8% of net generation at full load. Our experience has been favorable in this area, with the FGD system using slightly less than the predicted amount.

Landfill costs were predicted at $2.00 per ton of material. Our experience to date with the landfill operation has not met our expectations. The double alkali filter cake has generally been of a poorer quality than we anticipated, contributing to the problems and expense at the landfill. The two dollars per ton (1977 dollars) would have been adequate were it not for the difficulties resulting from the cake quality.

The double alkali system was purchased based on a lower evaluated cost. In cents per kilowatt hour, FMC's process was estimated to cost 0.269 vs. 0.306 for the next closest system, as reported by Wagner (2). Applying all appropriate correction factors to our experience, and inflating the other systems costs by actual reagent costs and the consumer price index for other items, the FMC process still exhibits the lowest cost for this installation, although not by as wide a margin.

CONCLUSION

In conclusion, SIGECO feels that the Double Alkali System installed at the A. B. Brown station can be successfully operated by utility personnel and can meet the requirements of Federal New Source Performance Standards while burning high sulfur midwestern coal.

REFERENCES

1. Durkin, T. H., Jaslovsky, G. S., and Boward, W. L., Operating Experiences with a Concentrated Double Alkali Process, American Power Conferences, April 23, 1980.

2. Wagner, N. P., Adams, L. J., and Ramirez, A. A., Technical and Economic Factors for Evaluating Flue Gas Desulfurization Technologies, American Power Conferences, April, 1978.

STATUS REPORT ON THE WELLMAN-LORD/ALLIED CHEMICAL FLUE GAS DESULFURIZATION PLANT AT NORTHERN INDIANA PUBLIC SERVICE COMPANY'S DEAN H. MITCHELL STATION

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TRW Inc.
Research Triangle Park, North Carolina

ABSTRACT

The Northern Indiana Public Service Company and the U. S. Environmental Protection Agency entered into a cost-shared contract in June of 1972 for the design, construction, and operation of a regenerable flue gas desulfurization (FGD) demonstration plant. The system selected for the project was a combination of the Wellman-Lord SO_2 Recovery Process and the Allied Chemical SO_2 Reduction Process. The FGD plant was to be retrofitted to NIPSCO's 115 MW pulverized coal-fired Unit No. 11 at the Dean H. Mitchell Station in Gary, Indiana. NIPSCO entered into contracts with Davy Powergas, Inc., for the design and construction of the FGD plant and with Allied Chemical Corporation for operation of the plant.

The FGD plant acceptance test was successfully completed on September 14, 1977. The plant completed a two-year demonstration test period during which information was collected and reported regarding pollution control performance, secondary effects, economics, and reliability of the system. TRW, Inc. was the independent evaluator for the EPA through October, 1979. A follow-on EPA/NIPSCO contract of seven and one-half months has recently been completed. Operation of the plant continues.

PROCESS DESCRIPTION

Wellman-Lord SO_2 Recovery

The Wellman-Lord process consists of three major operating sections - SO_2 absorption, purge treatment and SO_2 regeneration.

In the SO_2 absorption section, residual fly ash in the flue gas is removed by water scrubbing. SO_2 is then removed from the flue gas by scrubbing with a solution of sodium sulfite. The chemicals contained in this solution remain completely dissolved throughout the absorber. Flue gas scrubbing with a clear solution, free from suspended solids, plugging and scaling, is a fundamental reason underlying the exceptional on-stream reliability experienced in the commercial operations of the Wellman-Lord process.

The purge treatment section selectively removes inactive oxidized sodium compounds from a sidestream of the absorbing solution and converts this material into a dry granular product which is marketed.

The third section of the Wellman-Lord process involves thermal regeneration of the absorbing solution to release the absorbed SO_2 as a concentrated gas stream and return of the reconstituted solution to the absorber.

The concentrated SO_2 gas may be converted to liquid SO_2, sulfuric acid or elemental sulfur. NIPSCO elected to use the Allied Chemical SO_2 Reduction Process to convert to elemental sulfur.

Allied Chemical SO_2 Reduction to Sulfur

Sulfur is recovered by Allied Chemical's SO_2 reduction process which consists of two principal operating sections.

In the primary reduction section, more than one-half of the entering SO_2 is converted to elemental sulfur. A key feature of this section is the effective control of chemical reactions between SO_2 and natural gas over a catalyst developed by Allied Chemical for this purpose. Heat generated by these chemical reactions is recovered and utilized to preheat the SO_2 gas stream entering this section.

Packed bed regenerative heaters provide a rugged and efficient means for achieving this heat exchanger function. The process gas flow through the regenerators is periodically reversed to alternately store and remove heat from the packing; hence, the overall section is thermally self-sustaining.

Automatic control of the flow reversing cycles and other process conditions achieves optimum performance in the system, with high sulfur recovery efficiency and reductant utilization at all operating rates.

The gas leaving the primary reactor system is cooled in a sulfur condenser, for condensation and recovery of sulfur product. The remaining gas, containing proper proportions of SO_2 and H_2S is processed through a Claus conversion system for recovery of additional sulfur product. The Claus system off-gas is incinerated and recycled to the Wellman-Lord SO_2 absorber (see Figure 1). Since startup in 1977, 5843 long tons of sulfur have been produced.

PROCESS CHEMISTRY

The Wellman-Lord process is based on the chemistry of the sodium sulfite/bisulfite system: flue gas containing SO_2 is scrubbed with a sodium sulfite solution which absorbs SO_2, converting sodium sulfite to sodium bisulfite:

$$\text{(a)} \quad SO_2 + Na_2SO_3 + H_2O \rightleftharpoons 2\ NaHSO_3$$

The sodium bisulfite solution is regenerated by thermal decomposition. Application of heat simply reverses equation (a):

$$\text{(b)} \quad 2\ NaHSO_3 \rightleftharpoons Na_2SO_3 + SO_2 + H_2O$$

The SO_2 is recovered in a concentrated stream.

The concentrated stream of SO_2 gas is then reduced to high purity elemental sulfur in the Allied Chemical Process. This conversion is carried out in two steps. In the first step, a portion of the SO_2 in the feed gas reacts with natural gas, yielding a mixture of elemental sulfur, hydrogen sulfide, carbon dioxide and water vapor:

$$\text{(c)} \quad 2CH_4 + 3SO_2 \rightleftharpoons S + 2H_2S + 2CO_2 + 2H_2O$$

In the second step, H_2S formed in the first step reacts with the remaining SO_2 yielding additional elemental sulfur and water vapor:

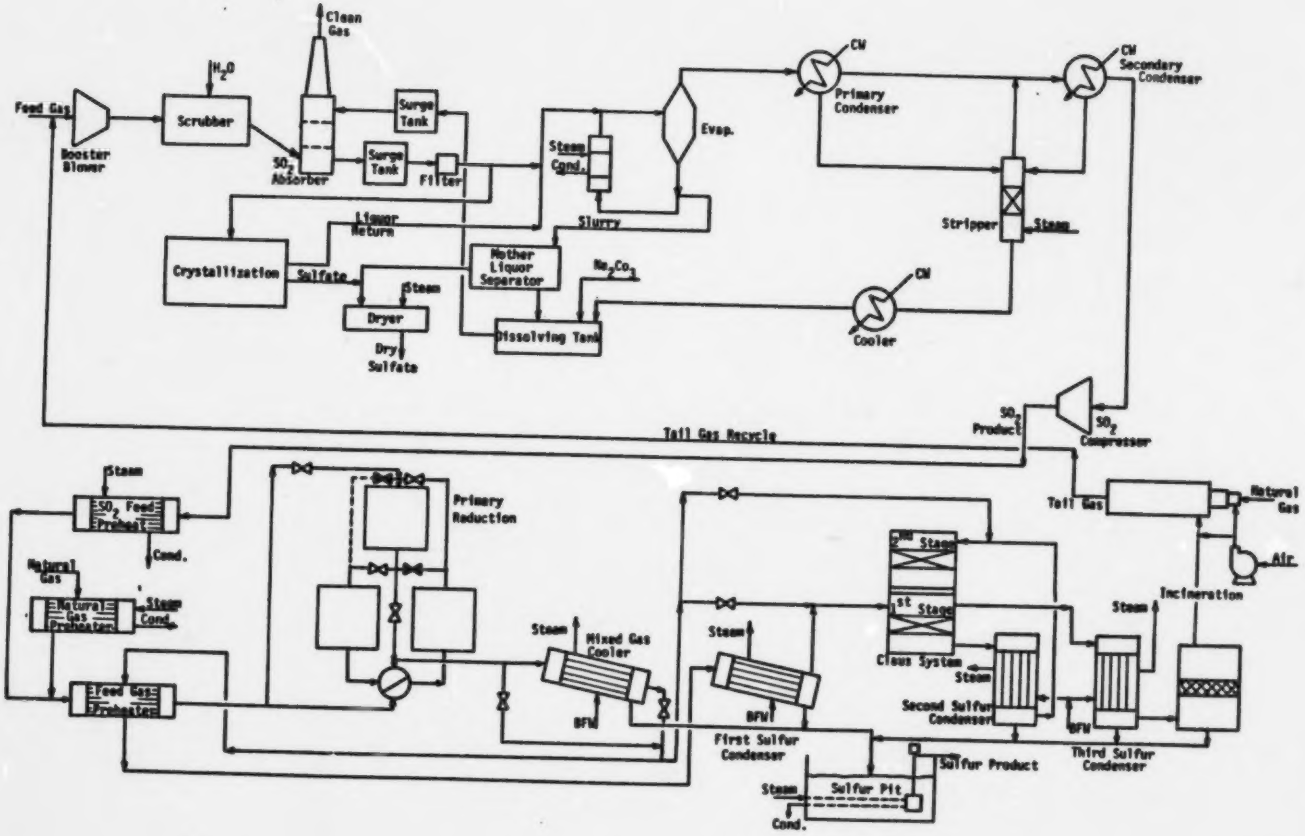

Feed Gas
Booster Blower
H_2O
Scrubber
Clean Gas
SO_2 Absorber
Surge Tank
Surge Tank
Filter
Steam
Cond.
Evap.
CW
Primary Condenser
CW
Secondary Condenser
Stripper
Steam
Liquor Return
Slurry
Crystallization
Sulfate
Mother Liquor Separator
Na_2CO_3
Steam
Dryer
Dry Sulfate
Dissolving Tank
CW
Cooler
SO_2 Product
SO_2 Compressor
Tail Gas Recycle
Steam
SO_2 Feed Preheat
Cond.
Natural Gas
Natural Gas Preheater
Steam
Cond.
Feed Gas Preheater
Primary Reduction
Steam
Mixed Gas Cooler
BFW
Steam
BFW
First Sulfur Condenser
2^{nd} Stage
1^{st} Stage
Claus System
Steam
Second Sulfur Condenser
Steam
BFW
Third Sulfur Condenser
Incineration
Tail Gas
Natural Gas
Air
Sulfur Product
Steam
Sulfur Pit
Cond.

Figure 1.
Wellman-Lord Recovery/Allied Chemical SO_2 Reduction
General Schematic Flow Diagram
NIPSCO Mitchell Unit No. 11

(d) $2H_2S + SO_2 \rightleftharpoons 3S + 2H_2O$

The tail gas from the sulfur plant is incinerated and recycled to the Wellman-Lord absorber.

TEST PROGRAM RESULTS

This section includes an analysis of the test results from the EPA evaluation that was conducted by TRW. The analysis focuses on the last thirteen months of the two-year demonstration period.

Description of Test Program

The test program as originally designed consisted of three major test phases:

(1) a baseline test

(2) acceptance testing

(3) a one year demonstration test and evaluation

The initial baseline and acceptance tests have been described in detail.[1,2] The acceptance test was successfully completed in September of 1977 and the scheduled one year of operation for demonstration testing followed immediately. During the demonstration year, operating experience was limited due to both boiler and FGD related operating problems. Operating experience and operating problems were described at the FGD Symposium held in March of 1979.[3] The test results have been reported.[4] These test results were inadequate for fully evaluating the FGD process because of those upsets caused by the boiler and thus external to the FGD plant. Modifications were begun during the latter half of the demonstration year that prompted the decision to continue with a demonstration test program for another full year. In this report, we will focus on the operating and SO_2 removal performance of the Wellman-Lord/Allied Chemical FGD unit during the second year of demonstration. The period covered is from October 1978 through October 1979. It was preceded by a second baseline test that provided up to date performance and operating data on the boiler while the FGD plant was down and completely isolated from the boiler.

As we stated earlier, modifications were begun during the latter half of the first year of demonstration. Except for insulation of the inlet ductwork, these modifications were completed during a scheduled boiler shutdown in September 1978. Our data show that the modifications ultimately enhanced the performance of the FGD unit and of the boiler. The boiler was utilized during 93% of the second year of the demonstration. With a dependable supply of flue gas to feed the FGD plant, conditions were quite favorable for gathering test data. The modifications that provided substantial improvement were as follows:

(1) Use of Captain coal. Coal feeding problems were minimized when this coal was used. Other corrective action for improving coal feeding were to enlarge the coal mill feed chutes and to increase capacity of the coal mills.

(2) Elimination of a boiler feedwater problem.

(3) NIPSCO agreed to remove a part of the heat transfer surfaces from the Ljungstrom air preheaters at some penalty in boiler efficiency. With this modification, flue gas temperatures were maintained above the dew point and booster blower problems caused by wet operation were eliminated.

(4) Electrification of the FGD evaporator circulating pump. Conversion from steam turbine to electrical drive reduced the startup time.

The test program demonstrates performance of the Wellman-Lord/Allied Chemical FGD process in these four major areas:

(1) Dependability of the FGD unit

(2) SO_2 removal performance

(3) Energy and raw material consumption

(4) Cost

The TRW test installation provided 10 or more measurements per hour of flue gas composition, steam and electrical energy consumption, and the boiler operating parameters of interest. One hour averages computed from these data served as the primary data base for most of the data interpretations. The amount of raw materials, natural gas and soda ash, and sulfur production were measured less frequently.

Summary of Results

The test program was designed to demonstrate guaranteed performance of the Wellman-Lord/Allied process and its ability to meet these performance criteria in a long term dependable manner and relative to the specific flue gas conditions at the host site. Since the FGD plant was designed and sized for a specific load factor and specific flue gas characteristics, the test also evaluated its operability over the normal range of load variation and flue gas composition experienced during the second demonstration year. The results are summarized as follows:

1. Reliability of the FGD unit, hours operated/hours called upon to operate, was 61%. The reliability record was established with virtually no redundancy built into the FGD unit. Also, the evaporator was designed for only 80% of full boiler load. With limited surge capacity within the regeneration loop, the FGD plant was not able to operate to effect complete SO_2 recovery during evaporator or reduction unit shutdowns.

2. The major sources of interruptions were
    - the reduction unit
    - the evaporator circulating pump
    - the booster blower
    - the evaporator
    - startup time

3. Twenty four-hour average SO_2 removal efficiences of 85% to 92% were typical. The pounds of SO_2 emitted per million Btu of heat input varied from 0.25 to 0.94.

4. SO_2 removal was attained at boiler loads in the range of 53 MWe to 85 MWe of the 115 MW boiler. Some operation was achieved up to 93 MWe. The lower limit was set by the limiting turndown capability of the reduction unit. The upper limit was set by

the capacity limitation of the evaporator as designed. This would not have been a limitation had the evaporator been designed to match the full SO_2 removal capability of the absorber. Since a substantial amount of energy largely as boiler main steam was consumed by the FGD plant, the generating potential of the boiler was actually about 95 MW at the FGD maximum capacity limit of 85 MWe.

5. SO_2 removal was attained from flue gas with the following characteristics relative to design:

   - SO_2 feed in excess of the expected plant capacity of about 5400 lb/hr was successfully treated for sustained periods of 24 hours and greater without loss of SO_2 removal efficiency. Overall for the second demonstration year, SO_2 feed averaged 4700 lb/hr.

   - Flue gas flow rates were usually higher than the expected flow rate of 320,000 acfm by a substantial amount. All flue gas flow rates in this report are at the design basis of $300^{o}F$ and one atmosphere.

   - Inlet temperature, following modifications to the air preheater of the boiler to obtain higher temperatures, averaged $305^{o}F$. Design basis temperature was $300^{o}F$.

6. The steam consumed by the FGD plant amounted to about 11% of the boiler input energy. Boiler derating averaged 9%.

7. Raw material consumption was as follows:

   - Soda ash average daily consumption was 9.9 tons. Moles of sodium consumed averaged 10.6% of the moles of SO_2 removed from the flue gas.

   - Natural gas was consumed at a rate of 7.1 cubic feet per pound of sulfur produced.

8. The production of sulfur as a byproduct averaged 17.1 tons per day of full operation. The product was sold locally.

FGD Dependability

There were various modes of operation depending on the availability of equipment and on flue gas availability. The principal operating modes were as follows:

Mode 1 - boiler operating, FGD not operating.

Mode 2 - integrated operation of the absorber/evaporator and reduction units without bypass of any of the flue gas.

Mode 3 - operation of absorber/evaporator with the flue gas bypass damper open. Bypassing of the flue gas may or may not occur depending on the booster blower speed setting.

Mode 4 - operation of absorber/evaporator with the reduction unit not operating. Recovered SO_2 (from evaporator overhead) is vented after dilution with flue gas from other units.

The FGD unit was considered to be fully operable only during Mode 2, although that also may have been the case during some of the Mode 3 operation. The exact operating status was difficult to determine while the bypass damper was open. However, failure to include any part of Mode 3 operation as fully operable time did not penalize the FGD process unfairly because the amount of accumulated full operation time with the bypass damper open was very low. Mode 2 or full operation status does not take into account the operation or performance of the purge treatment unit. The purge unit may or may not have been operating during Mode 2 operation and problems with the purge unit will be discussed later.

Figure 2 shows the reliability, hours of Mode 2 operation/hours called upon to operate, for the thirteen months of the second demonstration year. Called upon hours are those boiler operating hours when the boiler is delivering flue gas and steam within the design range. Figure 2 shows the FGD unit reliability factors plotted for each of the thirteen months. The overall average reliability was 61%. The ups and downs of operating performance shown here may be summarized as follows:

- Best reliability was achieved during October and November, 1978. For a 57 day period, October 16 to December 11, interruptions were minor and shortlived. FGD reliabiltiy was 99% during November.

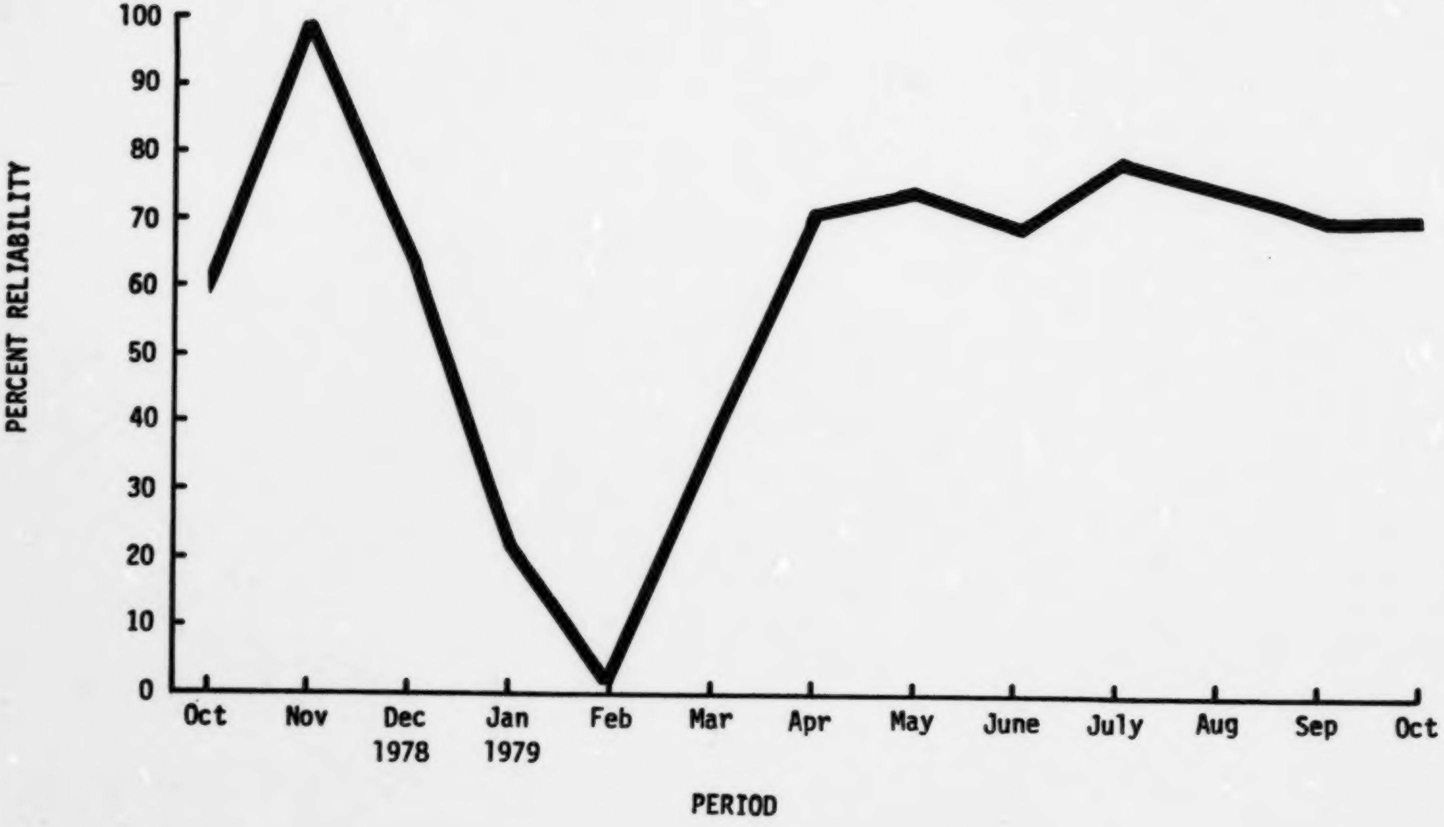


Figure 2. FGD Reliability Index

- Operation was limited to 66% reliability during December to clean the evaporator heater and for reduction area repairs.
- The FGD plant went down on January 10 for 43 days to repack the evaporator circulating pump and to retube a sulfur condenser.
- Full operation was limited during late February and early March due to numerous mechanical problems and leaks. The major problem during March was an outage of 16 days to overhaul and realign the evaporator circulating pump.
- From April until October, FGD plant reliability averaged 73%. Recurring problems with the booster blower turbine speed control and with the reduction unit were the primary limitations to better reliability.

It should be noted that during the seven month period, April to October, month by month reliabilities were fairly consistent, were primarily in a range of 70-75%, and were 13% higher than the thirteen month average reliability.

Table 1 identifies the equipment items that gave the most problems. The highest percentage of downtime was due to numerous interruptions of the SO_2 reduction unit. Since surge capacity for the scrubbing solution was minimal, any interruption of reduction unit operation required that either the evaporator be shutdown or recovered SO_2 be vented. Usual practice was to vent the SO_2.

The major interruption of the reduction unit was a 35 day shutdown to retube a sulfur condenser. Without this interruption, the reduction unit limited full operation during 7% rather than 17% of the called upon hours. The most frequent outages were those due to sulfur deposition, leaks, and valve repairs.

With electric instead of steam turbine drive for the evaporator circulating pump, emergency shutdown of the FGD plant was accomplished several times without difficulty. There were eight boiler shutdowns while the FGD plant was at full operation. Four of the shutdowns were without warning. Two of the shutdowns were after a short warning period of less than one hour. The other two occurred with adequate advance notice and the FGD shutdowns preceded the boiler shutdowns by 6 hours and 9 hours.

Table 1. REASONS FOR INTERRUPTION TO OPERATIONS

Equipment or reasons	Days of interruption	% of called upon time
Reduction unit	59	17
Evaporator circulating pump	28	8
Booster blower	18	5
Startup and shutdown	17	5
Other equipment, including absorber	10	3
Evaporator	8	2

Startups were a different story. Table 2 shows the startup time required for the same eight boiler shutdowns. Usual sequence after boiler startup was to start the absorber/evaporator loop and flue gas flow first followed by the reduction unit after which the bypass damper was closed. The startup record of Table 2 indicates, perhaps, that more time is required for startup after the more lengthy shutdowns. Otherwise, some of the startups seem to be unnecessarily long.

The other reasons for FGD plant interruptions may be summarized as follows:

- The evaporator circulating pump was down three times for repacking, for overhaul and for realignment of the motor shaft. It was also down once to replace a seal attributable to interruption of the steam supply from the boiler. Without steam, the condensate used for seal water was lost.
- The booster blower was down for relatively short periods but frequently. Most of the problems centered around the turbine governor and the gear reducer. There were no problems associated with the internal surfaces of the fan itself.
- There was one interruption for cleaning the evaporator heater after it had plugged.
- The absorber operated essentially trouble free. There was only one six hour interruption caused by an obstruction in the process water valve.
- Other problems accounted for less than 3% of the called upon time. They include frequent replacement of the SO_2 superheater with an overhauled spare and repairs to the SO_2 compressor. These items of equipment are in the line feeding SO_2 to the reduction unit. Other interruptions were for instrument and duct leak repairs.

SO_2 Removal

Removal efficiencies averaged 90% overall for the 13 months of operation. These removal efficiencies were obtained during an accumulated 211 days

Table 2. FGD PLANT STARTUP TIME

Boiler down hrs.	FGD startup time hrs	Startup sequence, hrs.		
		Absorber/evaporator	Reduction	Bypass damper
3	1	0	0.3	0.7
3	0.2	0	0.2	0
34	31	9	22	0.2
100	84	8	47	29
3	38	3	31	4
138	70	24	21	25
11	11	2	4	5
3	4	0	3	1
295	239	46	129	65

of Mode 2 operation of the FGD unit. Thirty day average removal efficiencies varied only from 88% to 93%. The range of values increases as averaging times decrease. Figures 3 and 4 are frequency distributions of 24-hour averages and one-hour averages, respectively. These data show that SO_2 removal performance compared to design (90%) and to the operating control point of 89% was as follows:

	% of time	
SO_2 Removal	24-hr Avg.	1-hr Avg.
90% and greater	60	52
89% and greater	84	78
85% and greater	97	97

It is seen that removal efficiencies of less than 85% occurred infrequently.

One-hour averages were accumulated to determine removal performance at the longer averaging times. Figure 4 shows that SO_2 removal was less than the limit of 89% for 22% of the time. The plant operated at removal rates of 85% or better for 97% of the time. The absorber was operated to achieve about 89% or higher removal rather than 90% removal. This was because operating control for SO_2 removal was set to reduce the SO_2 concentration on a diluted basis by 90%. Since dilution amounted to about ten percent, primarily as moisture, the actual removal was about one percent less than the reduction in concentration. The SO_2 removal data presented in this report have been corrected for dilution.

The flue gas being treated had the following characteristics. The characteristics of primary interest are: SO_2 feed, flue gas flow rate and temperature.

$\underline{SO_2\text{ Feed}}$. Sulfur level in the as received coal averaged 3.09 wt%. Confidence limits for this mean value at the 95% level were 2.99 wt% to 3.19 wt%. A coal of 3.16 wt% sulfur was used for design of the FGD plant. The estimated range, based on a distribution of two standard deviations from the mean, was 2.6 wt% to 3.6 wt%.

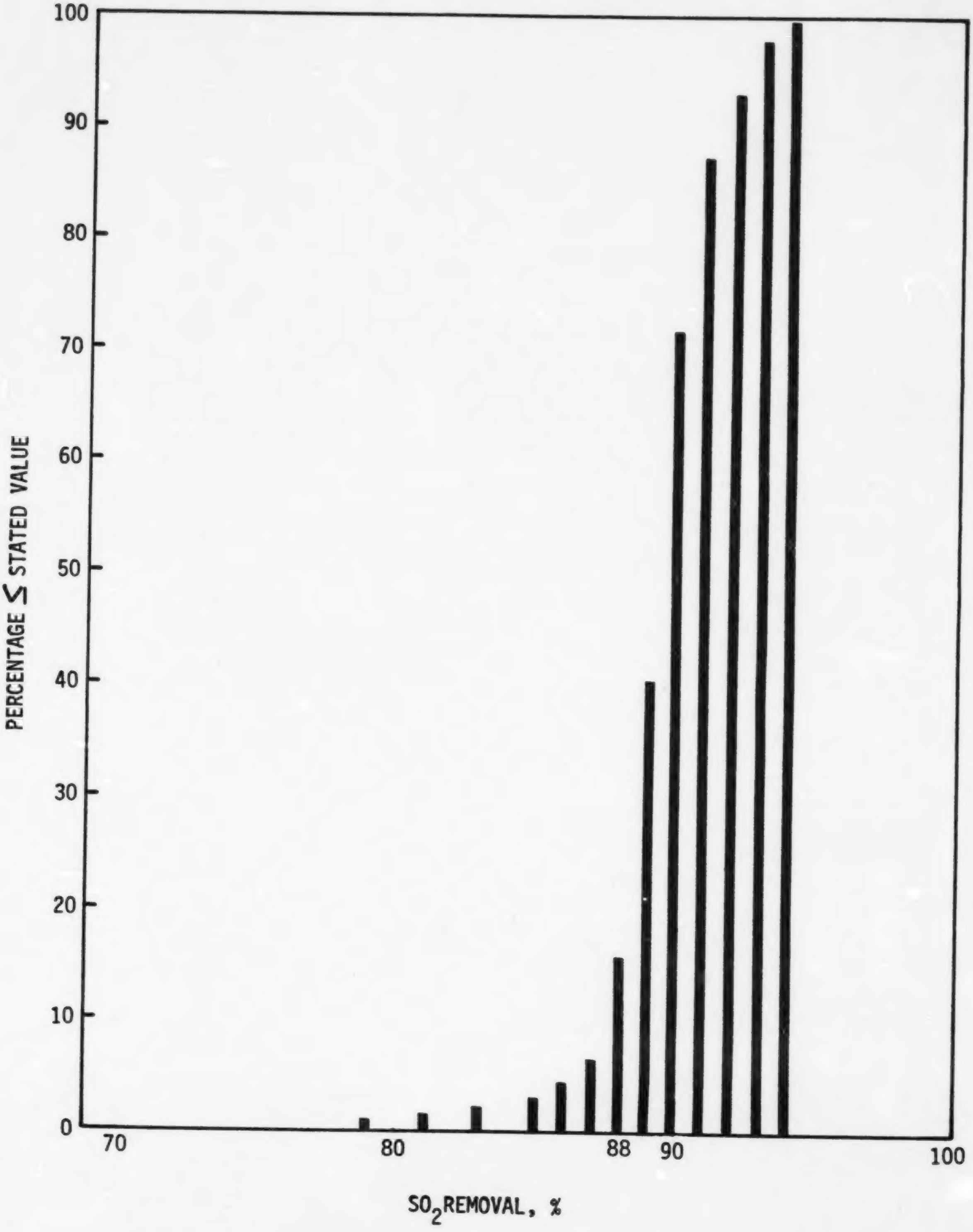


Figure 3. Cumulative Frequency Distribution of 24-Hour Average SO_2 Removal

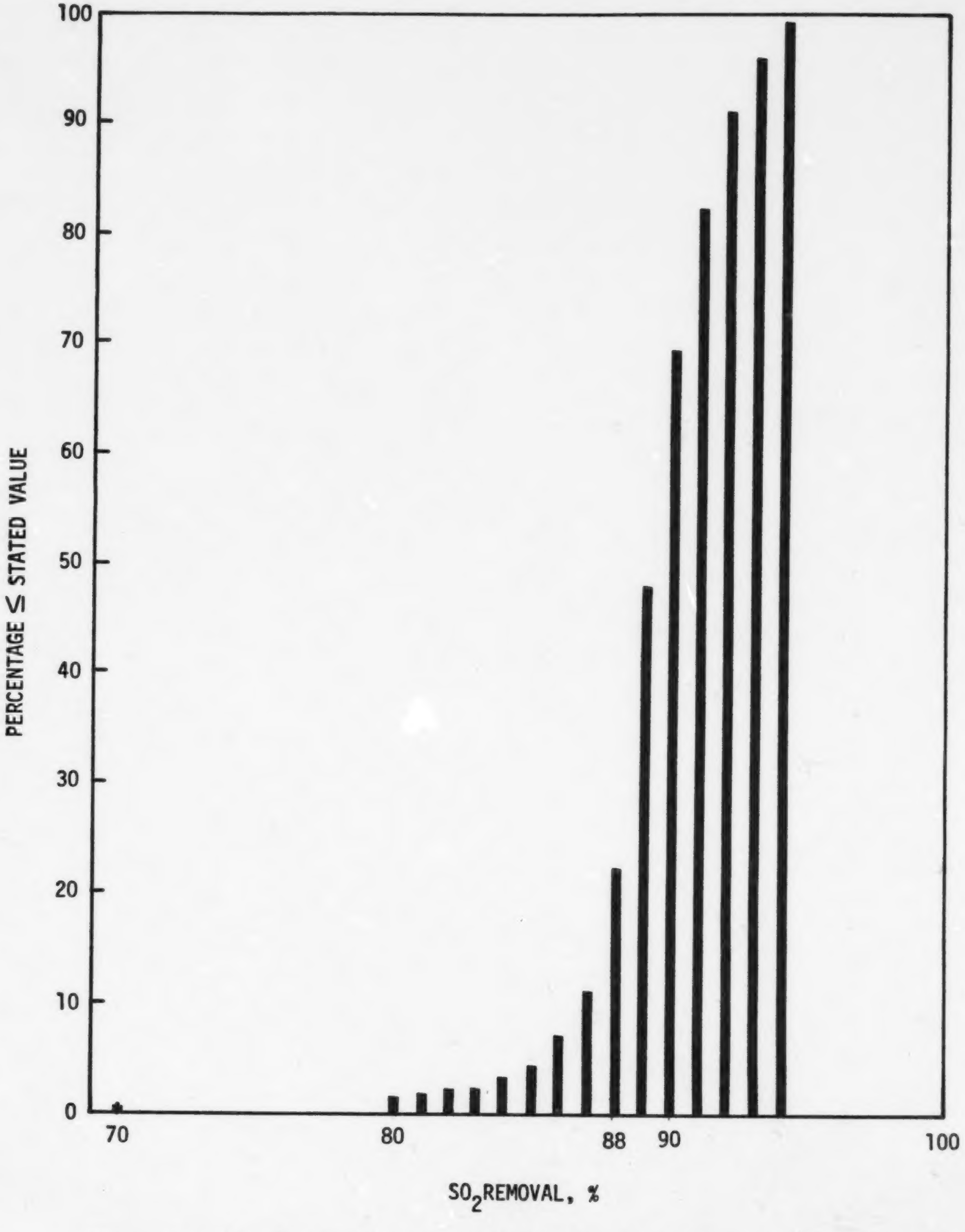


Figure 4. Cumulative Frequency Distribution of One Hour Average SO_2 Removal

Concentration of SO_2 was expected to be 2185 ppmv, wet basis, at a coal sulfur level of 3.16 wt%. The equivalent SO_2 feed rate for this concentration is 4842 lb/hr at the design flue gas flow rate of 320,000 acfm and the design equivalent boiler load factor of 80%. Actual flue gas flows were somewhat higher due to higher than expected excess air caused by excessive inleakage believed to be at the air preheaters. The absorber was designed to receive a volume of flue gas equivalent to 100% load factor and to remove the corresponding amount of SO_2 at inlet concentrations of at least 2185 ppmv. While the capacity to handle flue gas flow rates and SO_2 feed rates equivalent to 100% load factor was demonstrated for short periods, sustained operation was only possible at load factors slightly better than 80% because of the limited capacity of the evaporator and limited surge capacity in the regeneration loop. This limited the capacity of the boiler during high demand that was in addition to the derating effect from FGD plant steam consumption. While performance at high load was not fully demonstrable, minimum sustainable operating rates were demonstrated during turndown tests. Full operation at 53 MW was sustained for four days. Table 3 summarizes the inlet conditions of this test.

Table 3. FGD PLANT MINIMUM SUSTAINABLE LOAD TEST RESULTS

Length of test	4 days
Average load	53 MWe
Flue gas flow	237,000 acfm
SO_2 feed	3,381 lb/hr
SO_2 feed	2,069 ppmv wet
Oxygen in flue gas	7.5 vol. % dry

Below this operating level, the FGD plant is limited by the turndown capability of the reduction unit. The absorber/evaporator was operated down to 44 MWe. It must be remembered, however, that these minimum loads are generator output after derating due to the steam requirements of the FGD plant. Equivalent loads relative to flue gas flow and boiler heat input are 50 MW generating potential at the 44 MWe level and 60 MW generating potential at the 53 MWe level.

Boiler Load. The FGD plant operated with SO_2 removal rates of 89% or better at 24-hour average loads from 53 MWe to 93 MWe. However, 85 MWe was usually the upper limit for operation. Figure 5 presents the load factors for each 30-day period of the second demonstration year during Mode 2 operation of the FGD plant. The FGD plant was expected to operate at a boiler load potential of 92 MW and did indeed meet or exceed this capacity during three of the 30-day periods. For the rest of the time, the boiler was demand limited and the load factors remained below the 92 MW level.

Flue Gas Flow Rate. While boiler load was generally below the design level, the reverse was the case for flue gas flows. This was due to higher than expected excess air in the flue gas. Figure 6 shows flue gas flow rates as a function of both the actual generator output and the boiler load potential. Flue gas flows of 320,000 acfm, the design level, were attained at 71 MW of generator output or 80 MW of load potential. At a FGD load limit of 85 MW, the flue gas flow rate was over 360,000 acfm. At the load potential of 92 MW, the design point, the flue gas flow rate was nearly 360,000 acfm. Since the absorber was designed for full load, the greater volume of flue gas presented no apparent problems for the booster fan or the absorber.

Excess Air. The high flue gas flow rates are explained by the higher than expected amount of excess air that can be attributed primarily to inleakage air believed to be entering at the air preheaters. Oxygen levels in the flue gas averaged 8.0% by volume compared to an expected oxygen level of 5.6%. On average, the additional amount of air would increase the total quantity of flue gas by about 17%. At higher than average loads, the excess air averaged a little less than the overall average and would add a little less than 17% to the quantity of flue gas.

Temperature. Inlet temperatures averaged 305°F during the second demonstration year. However, 38% of the hourly average temperatures

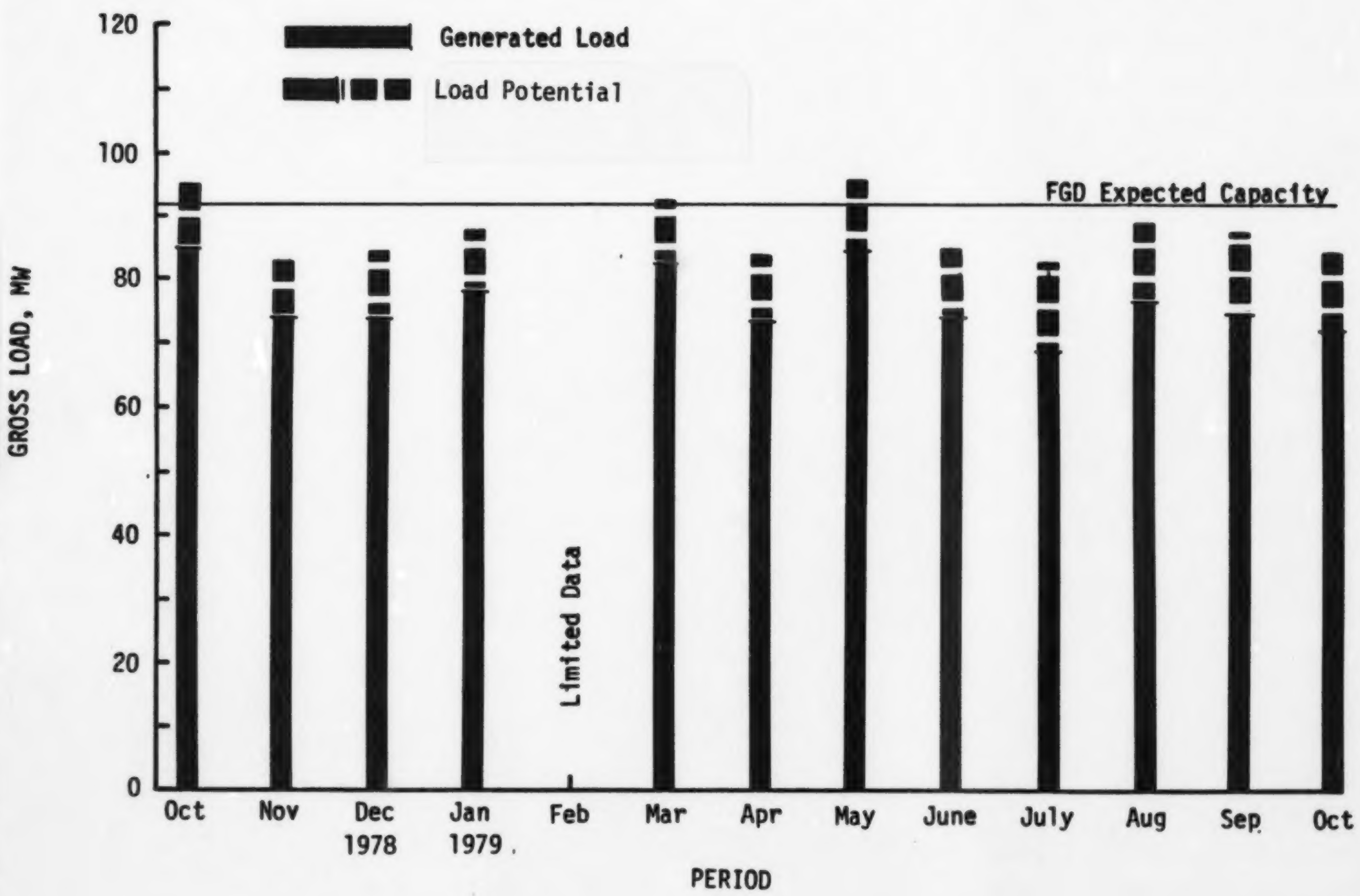


Figure 5. Boiler Load and Load Loss During FGD Operation

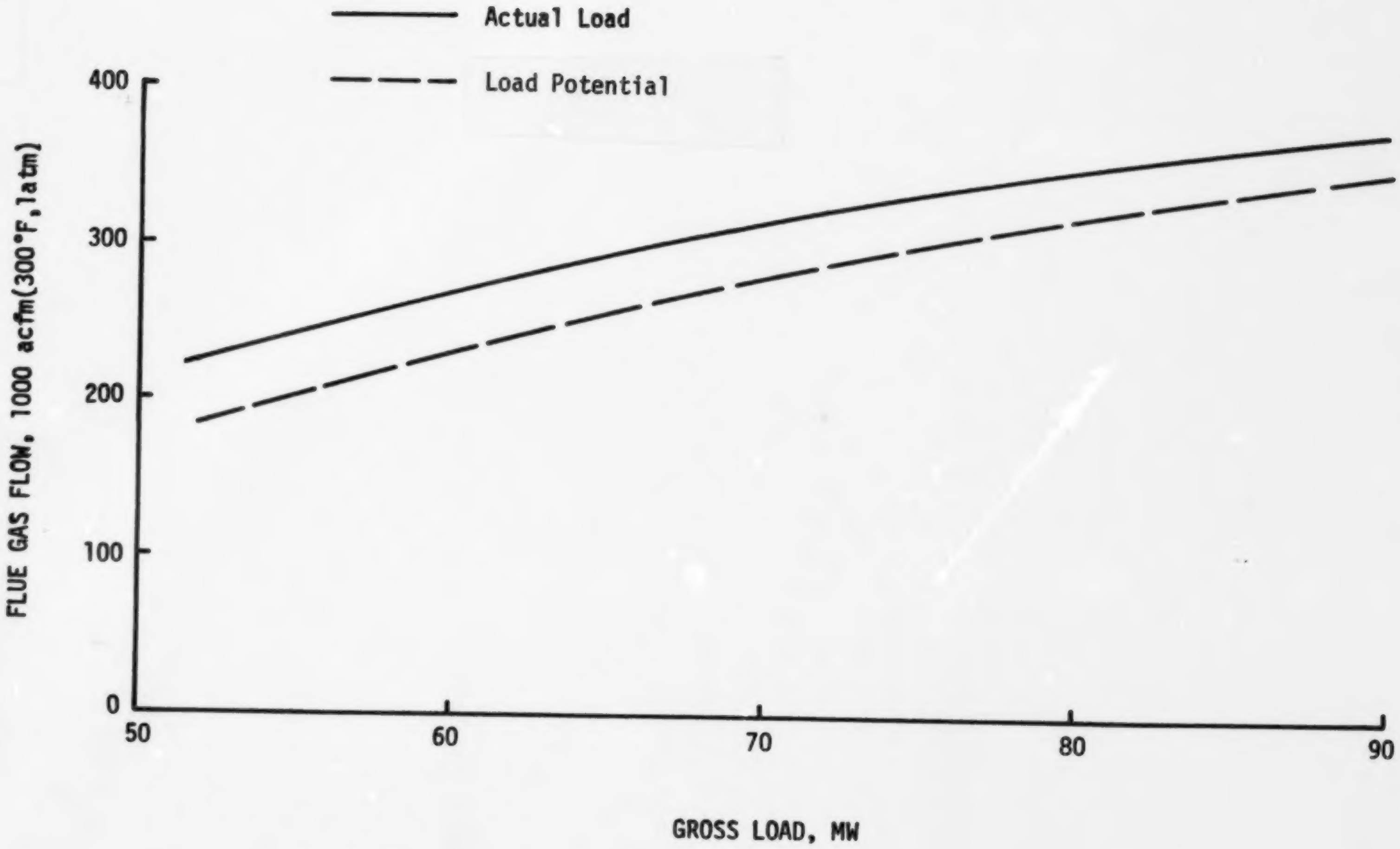


Figure 6. Flue Gas Flow Rates at Actual Load and Load Potential

were below 300°F but virtually all of these temperatures were above 280°F. It should be noted that these are single point temperatures well within the flue gas stream and do not reflect temperatures at and near the duct surfaces. There were, nevertheless, no problems attributable to wet flue gas.

Energy Consumption

A significant amount of the steam produced by the boiler was consumed by the FGD plant, primarily for operation of the evaporator for recovering the SO_2 and regenerating the scrubber solution. Boiler main steam from the superheater at 1800 psi and 1000°F was let down and desuperheated to obtain steam for the FGD plant at 550 psig and 750°F. This steam was used in steam turbines to drive the booster blower, SO_2 compressor and the evaporator circulating pump. However, before the start of the second demonstration year, the evaporator circulating pump drive was electrified to eliminate the startup and shutdown problems that occurred when high pressure steam was interrupted by unscheduled boiler shutdowns. The turbine exhaust steam along with additional 550 psig steam that had been let down through a pressure reducing valve was desuperheated further and used for process heat, primarily at the evaporator.

Actual steam consumption (at 550 psig, 750°F) varied from 52,000 lb/hr to 59,000 lb/hr during the second demonstration year. In Btu's, this was equivalent to 11% of the boiler input energy derived from fuel and derated the 115 MW boiler by 8% at the average boiler load of 77 MWe. In addition to steam consumption, about 700 kW of electricity was consumed, exclusive of the evaporator circulating pump motor. This increases the total energy requirement to about 12% of the boiler heat input derived from fuel. This derated the boiler another 0.6%. The total derating is equivalent to an electric production loss of 10 MW of generator output. Power to the evaporator circulating pump was not metered but is estimated at 330 kW.

Raw Material Consumption

Soda ash is used as makeup sodium carbonate for the scrubbing process. Usage is determined by buildup of inactive constituents in the absorber/

evaporator loop, such as sulfate and thiosulfate, that have to be purged. Any loss from the system due to leaks would also require soda ash makeup. High soda ash consumption during the first demonstration year were due to leaks at the bottom collector tray of the absorber that were repaired before start of the second demonstration year. These leaks effectively aborted the estimation of purge rates during the first year.

For the thirteen months of the second demonstration year, 2273 tons of soda ash were consumed, for an average daily consumption of 9.4 tons per day, using the total operating days of the absorber/evaporator as the time base. The performance guarantee for acceptance was 6.6 tons per day at the design levels of flue gas flow and inlet SO_2.

Natural gas is used as the reductant for converting the SO_2 to elemental sulfur. It is also the fuel used to incinerate the tail gas emitted from the reduction process. The tail gas is returned to the inlet of the absorber after incineration. It was necessary that the incinerator continue to be operated during shutdowns for destruction of the reduced sulfur forms that desorb from the reduction unit refractory materials. Thus, there is a corresponding improvement in unit consumption of natural gas with improvement in reliability. Table 4 shows that slightly over 7 cubic feet of natural gas was consumed per pound of sulfur produced.

Table 4. NATURAL GAS CONSUMPTION

Annual consumption, million cf	54.1
For process use, %	87.9
For incineration use, %	12.1
Average consumption during mode 2 operation, cf/hr	9745
For process use, %	92.5
For incinerator use, %	7.5
Process gas/sulfur produced, cf/lb	6.2

(continued)

Table 4 (continued)

Total gas/sulfur produced, cf/lb	7.1
Consumption during shutdown, % of total	5.6

This meets the design expectations. Average consumption was 9745 cf/hr. of which 7.5% was consumed by the incinerator. In contrast, the incinerator consumed over 12% of the total gas overall for the second demonstration year and is a consequence of the 61% FGD plant reliability factor.

Purge Treatment Limitations

The purge unit as initially designed was to have treated a small purge stream removed from the regeneration loop, to effect a separation of sodium sulfate from most of the sulfite/bisulfite components, and to dry the sodium sulfate to produce a salable product. The "wet" end of this purge treatment system performed satisfactorially but the dryer had a capacity of only about 50% of that needed. The requirements on the purge unit and the drying problem will be discussed in turn.

The amount of purge to be treated is a function of the formation of sulfate and possibly thiosulfate during absorption. Attempts were made by TRW to determine the amount of sulfate formation during absorption but these efforts were frustrated by inability to obtain correct flow measurements and uncertainties about the specific water balance across the absorber. However, the data seem to indicate that sulfate formation is a function of oxygen concentration in the flue gas. Since excess air levels were higher than design expectations, higher than design purge rates might be necessary. Purge rates were not measured directly. However, an average purge rate for the period April through October has been estimated at between 10.6 and 12.4%. Purge rate is the ratio of moles sodium in purge to moles SO_2 removed from flue gas, expressed as a percentage. The estimate was determined from soda ash consumption and the calculated amount of SO_2 removed. A purge rate of about 10% was the value indicated during the design phase of the project. In the aggregate, the above information seems to point to actual purge rates higher than design, the magnitude of which is unclear. As stated before, the process up to drying seemed to perform satisfactorially.

Dryer tests performed by Davy McKee determined that the dryer would not work, even at design rates. There had always been a question of whether the sulfate dryer actually had heat duty design capacity. In May of 1979, tests verified that the dryer did not have design capacity. The purpose of the test was to demonstrate the heat capacity of the dryer with a water feed onto a sodium sulfate bed in the dryer. If the maximum rate could be reached with water, the dryer was then to be tested with the sodium sulfate solution recovered from the purge solution.

The maximum dryer capacity achieved during the test was approximately 66% of the design heat duty. Capacities of 59-66% of design were maintained for 2-1/2 days. However, after operating for 2-1/2 days at 59-66%, the motor tripped out several times because of an amperage overload. When the motor tripped out, a buildup of solids was observed at the discharge end of the dryer. To prevent overloading the motor, the water feed rate was reduced, so during the last 2-1/2 days of the test, the dryer capacity dropped to 45-50% of design. At this point, further testing was abandoned.

Possible solutions include:

1. A more concentrated feed fed to same dryer.
2. A dryer of different design.
3. Addition of an antioxidant to the absorbing solution.

Replacement of the equipment capable of attaining design capacity would cost approximately $500,000. Alternates to the equipment replacement were sought. Antioxidants to reduce sulfate formation were considered. Tests of EDTA (Ethylene Diamine Tetra Acetic Acid) were run April - May of 1980. The tests were terminated after five days of an intended two-week test because of unrelated equipment failure. The too brief test period indicated a possible 50% reduction in sulfate formation. Additional testing of EDTA is planned.

OPERATING AND MAINTENANCE COSTS

NIPSCO UNIT NO. 11 FGD PLANT

Operating, Maintenance and Improvement Costs are listed from the period of January 1, 1979, through April 30, 1980.

Operation and Maintenance - Offsites Facilities (including booster blower, flue gas ductwork and dampers, utilities system)	$ 154,160
Operation and Maintenance - FGD Process (includes by-products storage and loading, raw materials unloading and storage, and Allied Management Fee)	6,061,205
Total FGD Costs before By-Product Credit	6,215,365
Credit for By-Products	104,963
Total FGD Operation/Maintenance Costs after By-Product Credit	$6,110,402

No amortization costs are included above.

COAL DATA AND UTILITIES - January 1, 1979, through April, 1980

High Sulfur Coal Burned, Tons	349,121
Average BTU/#	10,586
Average Sulfur %	2.85
Steam Used, Pounds	455,000,000
Boiler Feedwater Used, Pounds	29,128,000
Condensate Returned to Gen. Sta., Pounds	333,120,000
Condensate Dumped, Pounds	151,008,000
Electric Power, Kwh (Including 500 HP Circ. Pump Motor)	6,472,000
Natural Gas, Ft^3	57,642,000
Service Water, 1000 Gallons	2,784,000
Elemental Sulfur Sold, Long Tons	2,668
Sulfate Sold, net tons (no dry sulfate was produced in 1980)	139.6

NORTHERN INDIANA PUBLIC SERVICE COMPANY
DEAN H. MITCHELL STATION
UNIT NO. 11 SO_2 DEMONSTRATION PLANT
OPERATING HOURS GRAPH

1. Solid line indicates Unit No. 11 or SO_2 Plant is in operation.

2. Definition of Unit No. 11 being in operation is: Unit synchronized on line regardless of megawatt load.

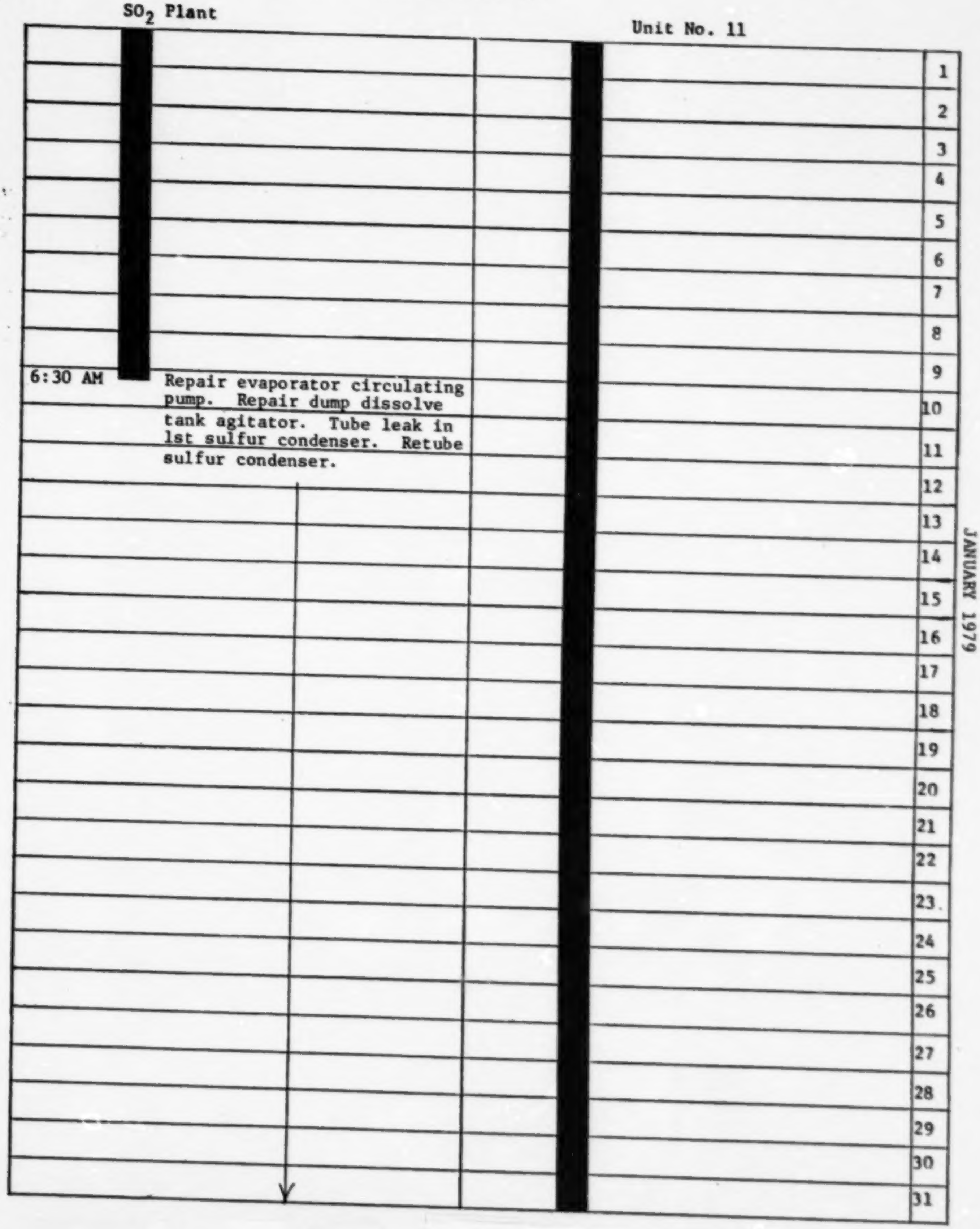
3. Definition of SO_2 Plant being in operation is:

   a. Receiving all of flue gas from Unit No. 11.
   b. No SO_2 bypassed to the Unit 6-11 stack.

4. Unit 11 operating conditions required for SO_2 Plant Operation are:

   a. Unit 11 operating on high sulfur coal at 46 GMWE (min.).
   b. Sufficient main steam available (530 PSIG minimum).
   c. Sufficient demineralized make-up water available.
   d. Unit 11 supplied utilities available (electricity, boiler feed water).

OPERATING HOURS

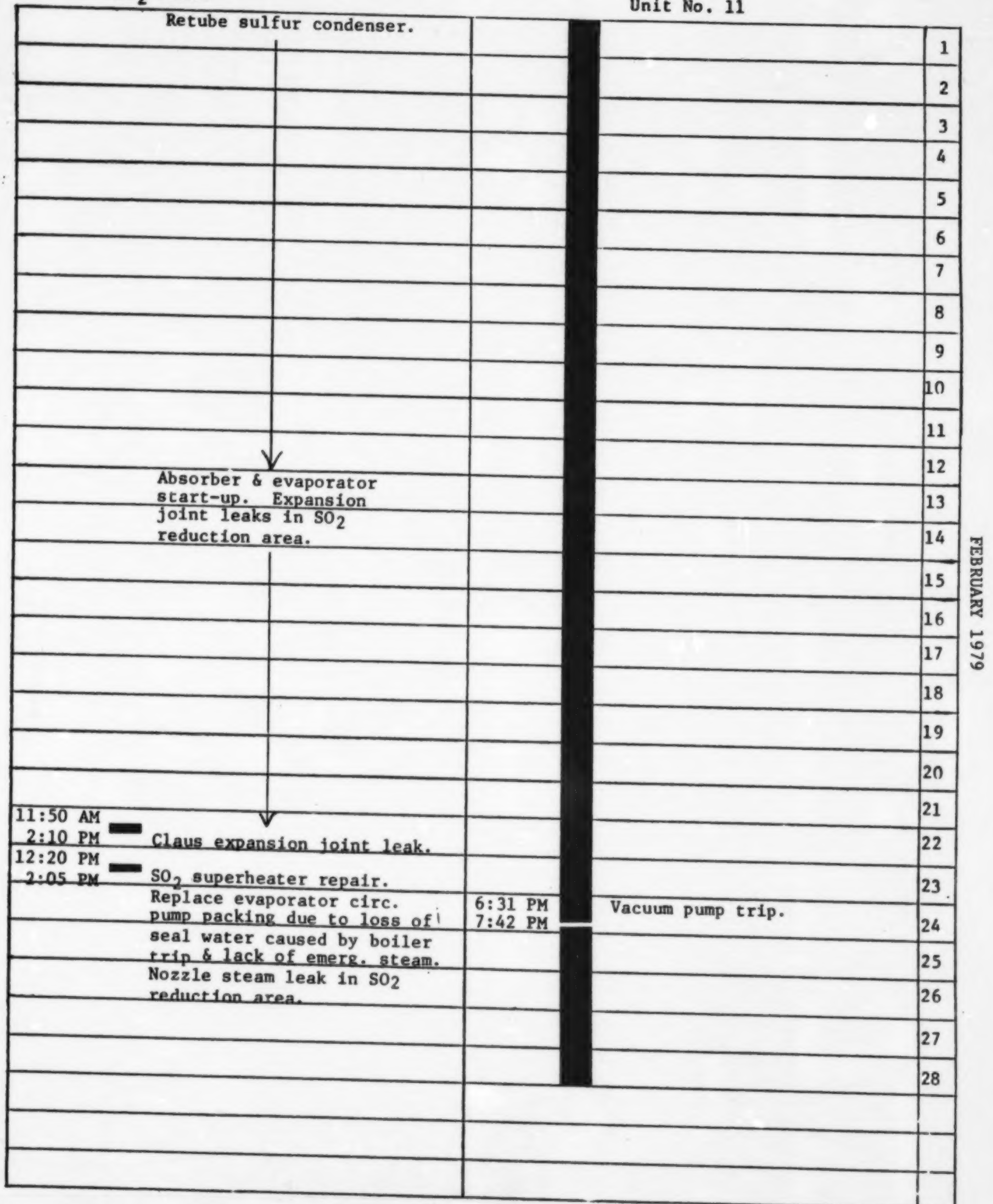


6:30 AM Repair evaporator circulating pump. Repair dump dissolve tank agitator. Tube leak in 1st sulfur condenser. Retube sulfur condenser.

OPERATING HOURS

SO$_2$ Plant

Unit No. 11



Retube sulfur condenser.

Absorber & evaporator start-up. Expansion joint leaks in SO$_2$ reduction area.

11:50 AM
2:10 PM

Claus expansion joint leak.

12:20 PM
2:05 PM

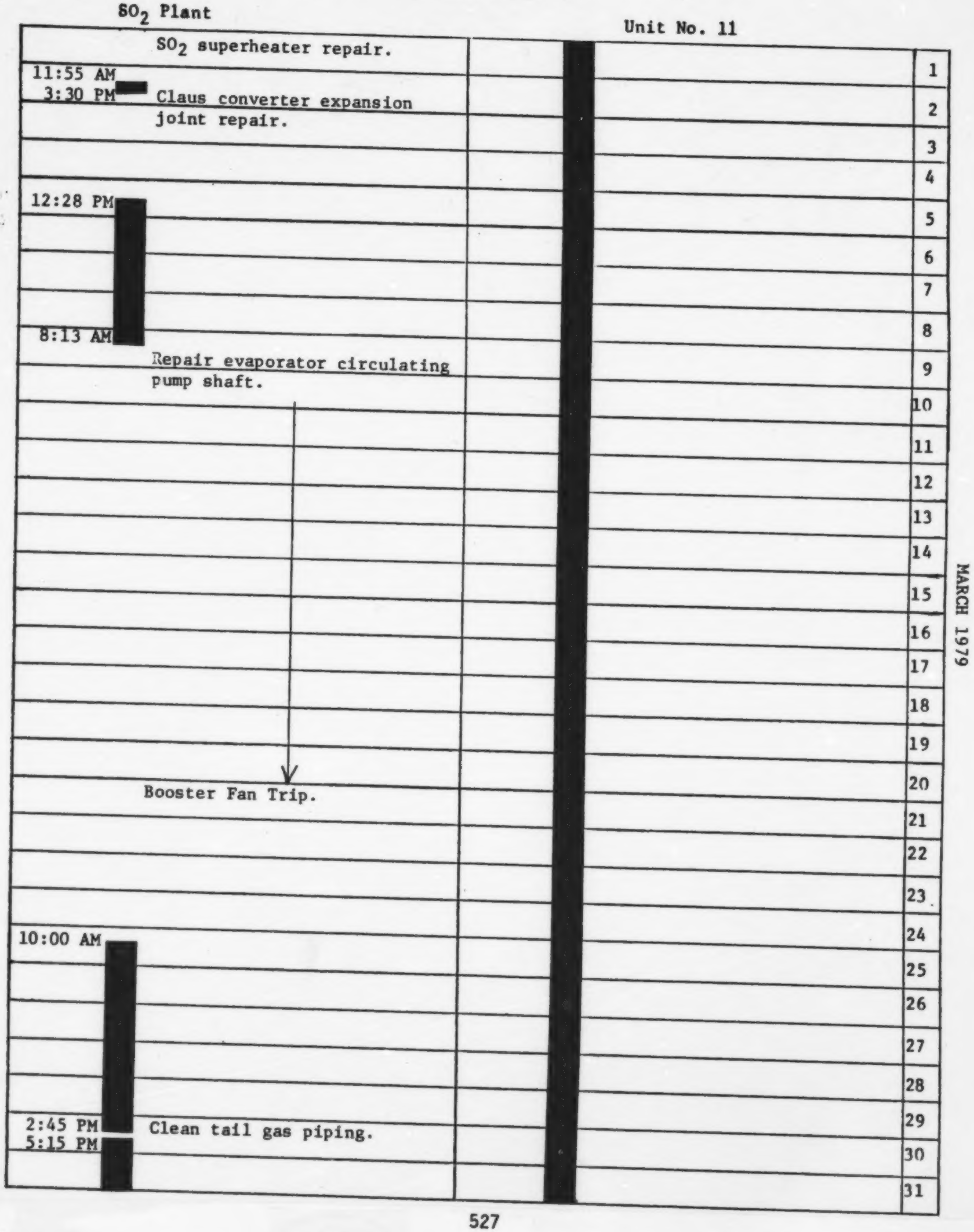
SO$_2$ superheater repair. Replace evaporator circ. pump packing due to loss of seal water caused by boiler trip & lack of emerg. steam. Nozzle steam leak in SO$_2$ reduction area.

6:31 PM
7:42 PM

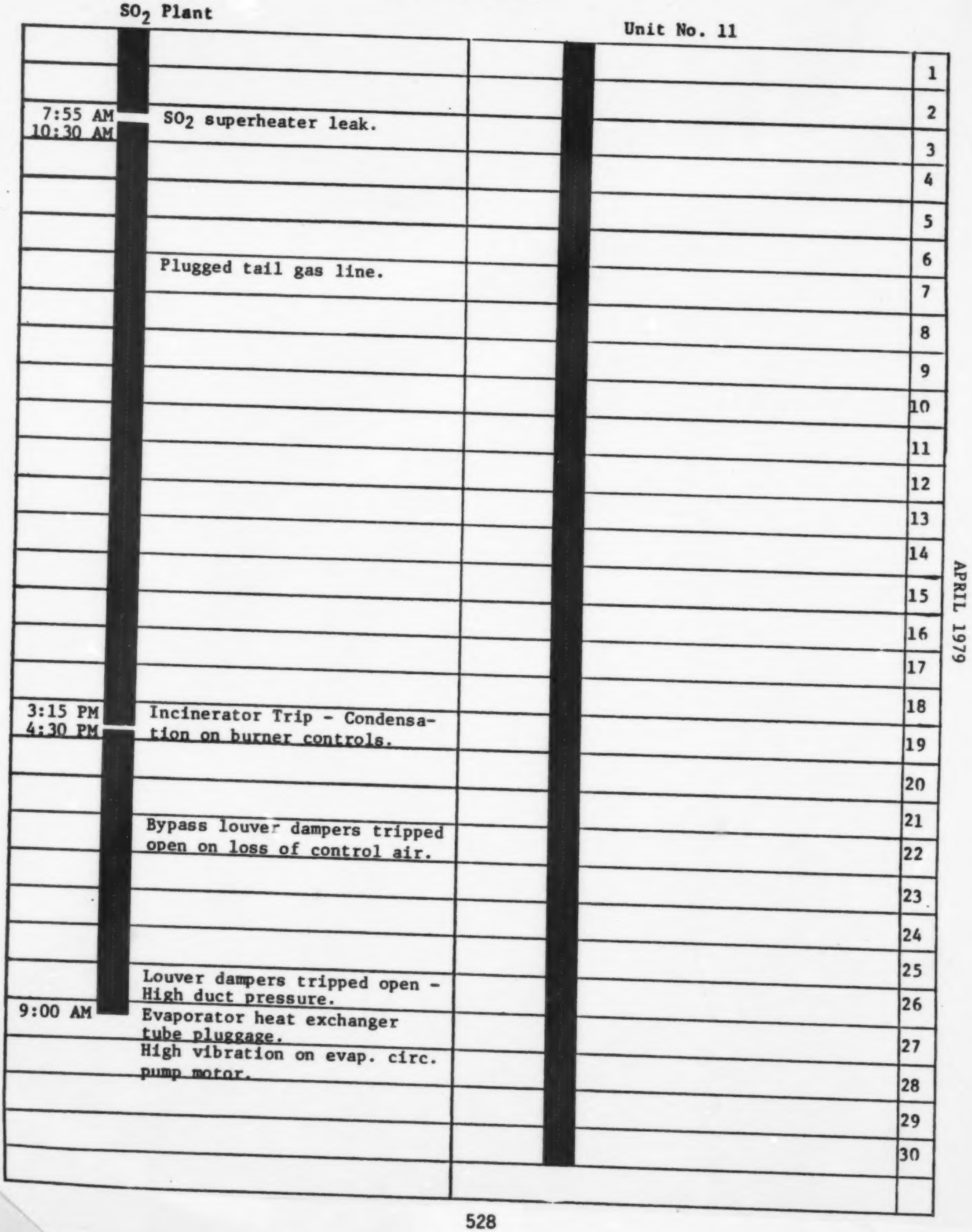
Vacuum pump trip.

FEBRUARY 1979

OPERATING HOURS



OPERATING HOURS



7:55 AM
10:30 AM
SO$_2$ superheater leak.

Plugged tail gas line.

3:15 PM
4:30 PM
Incinerator Trip - Condensation on burner controls.

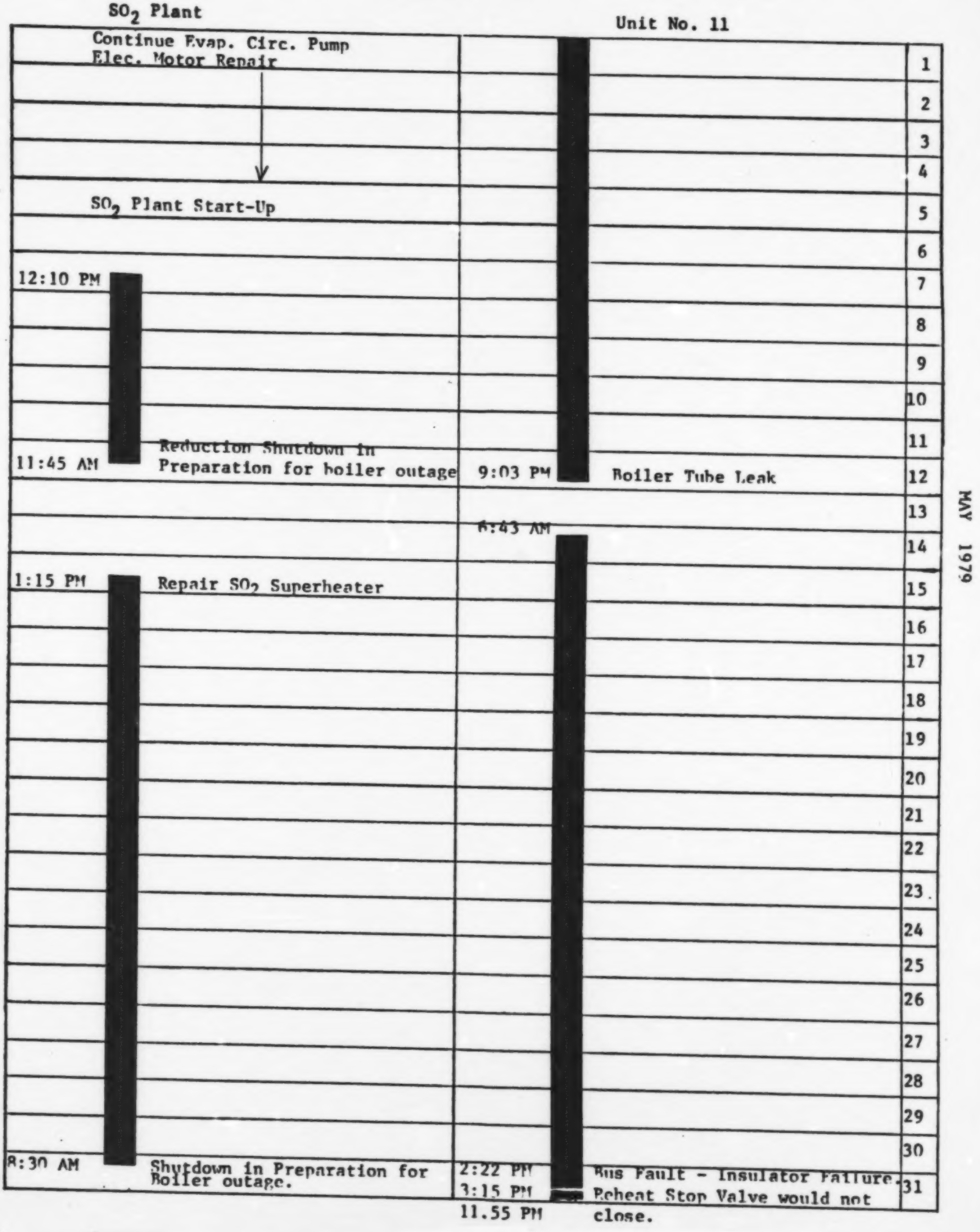
Bypass louver dampers tripped open on loss of control air.

Louver dampers tripped open - High duct pressure.

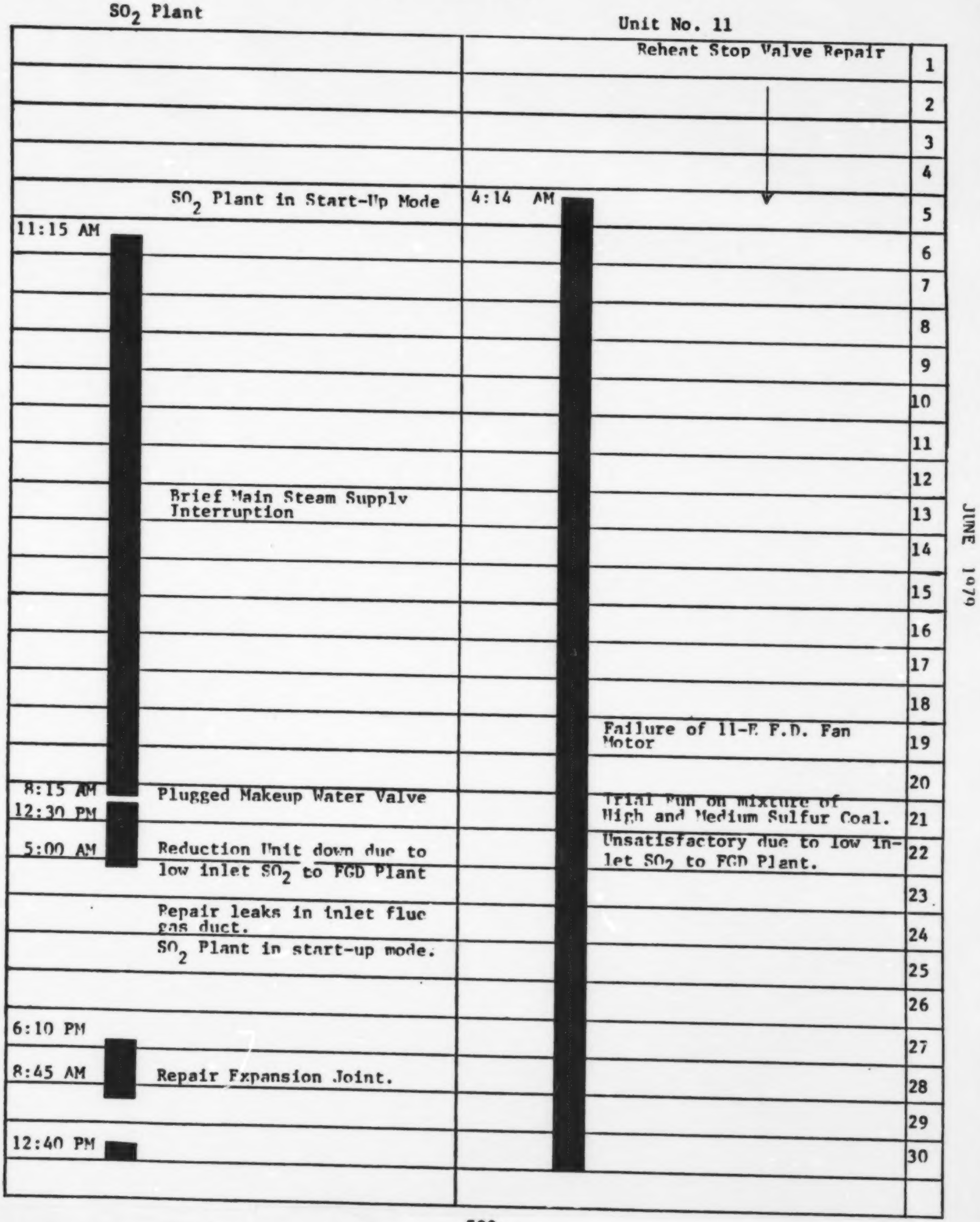
9:00 AM
Evaporator heat exchanger tube pluggage.

High vibration on evap. circ. pump motor.

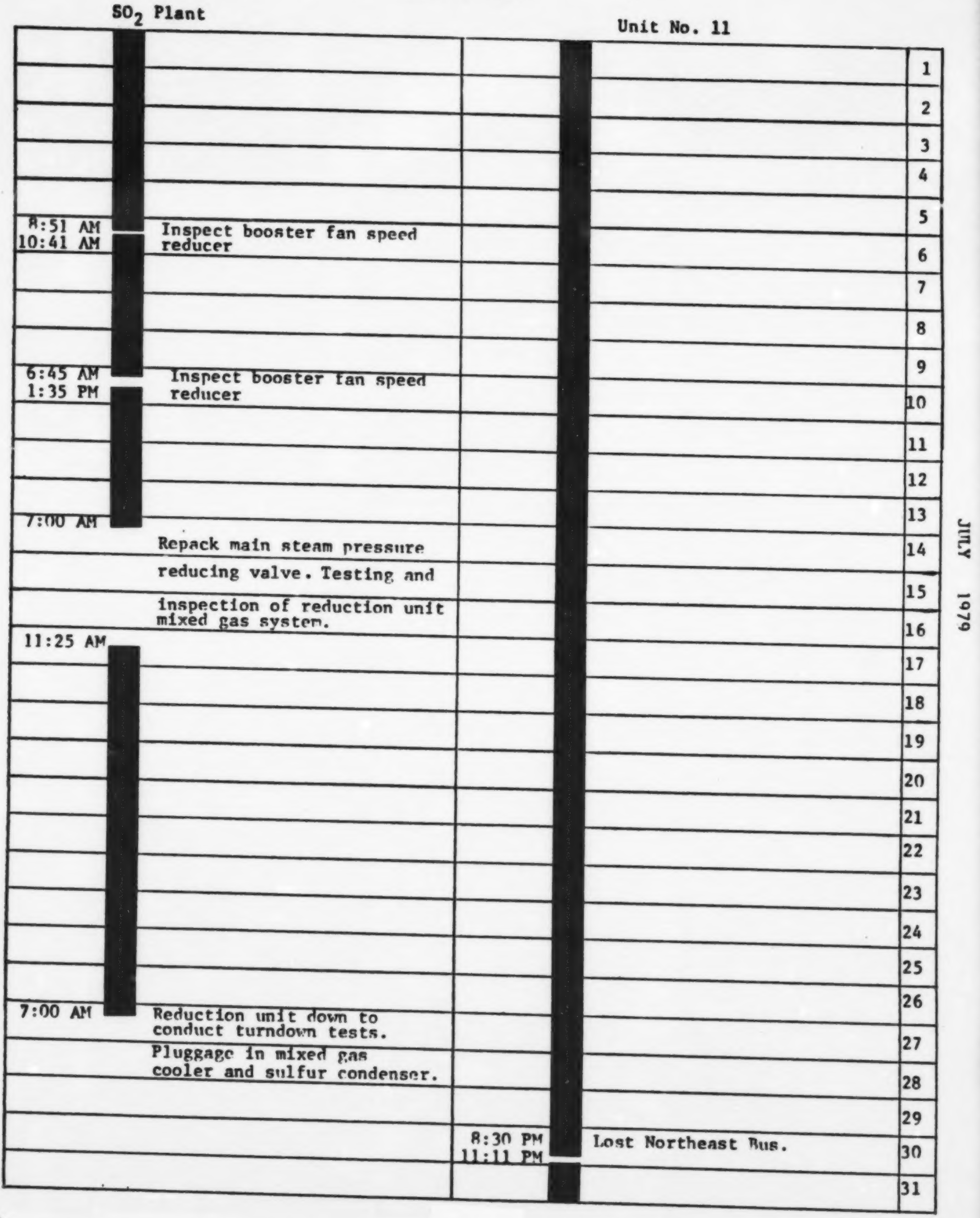
OPERATING HOURS



OPERATING HOURS



OPERATING HOURS



8:51 AM
10:41 AM
Inspect booster fan speed reducer

6:45 AM
1:35 PM
Inspect booster fan speed reducer

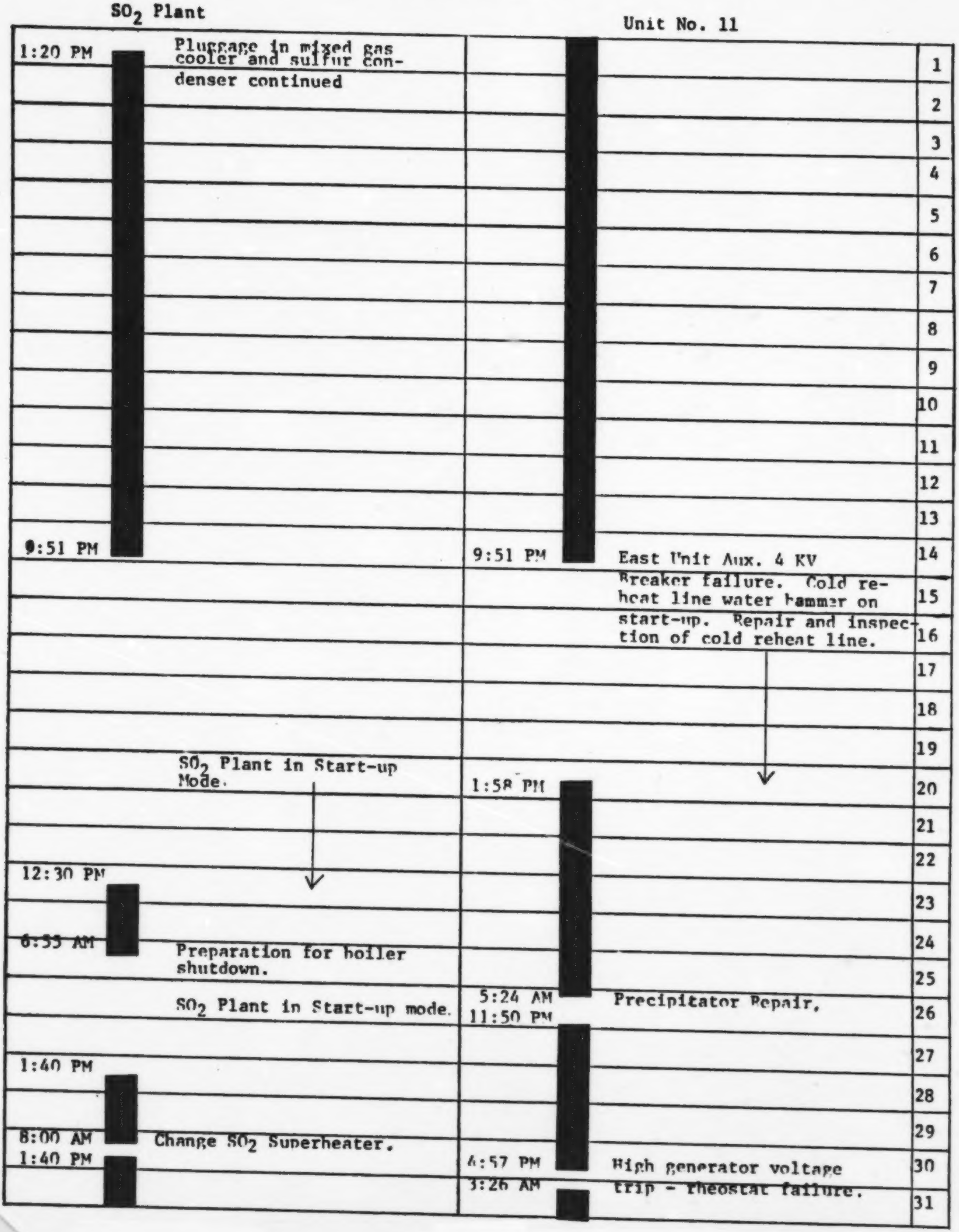
7:00 AM
Repack main steam pressure reducing valve. Testing and inspection of reduction unit mixed gas system.

11:25 AM

7:00 AM
Reduction unit down to conduct turndown tests. Pluggage in mixed gas cooler and sulfur condenser.

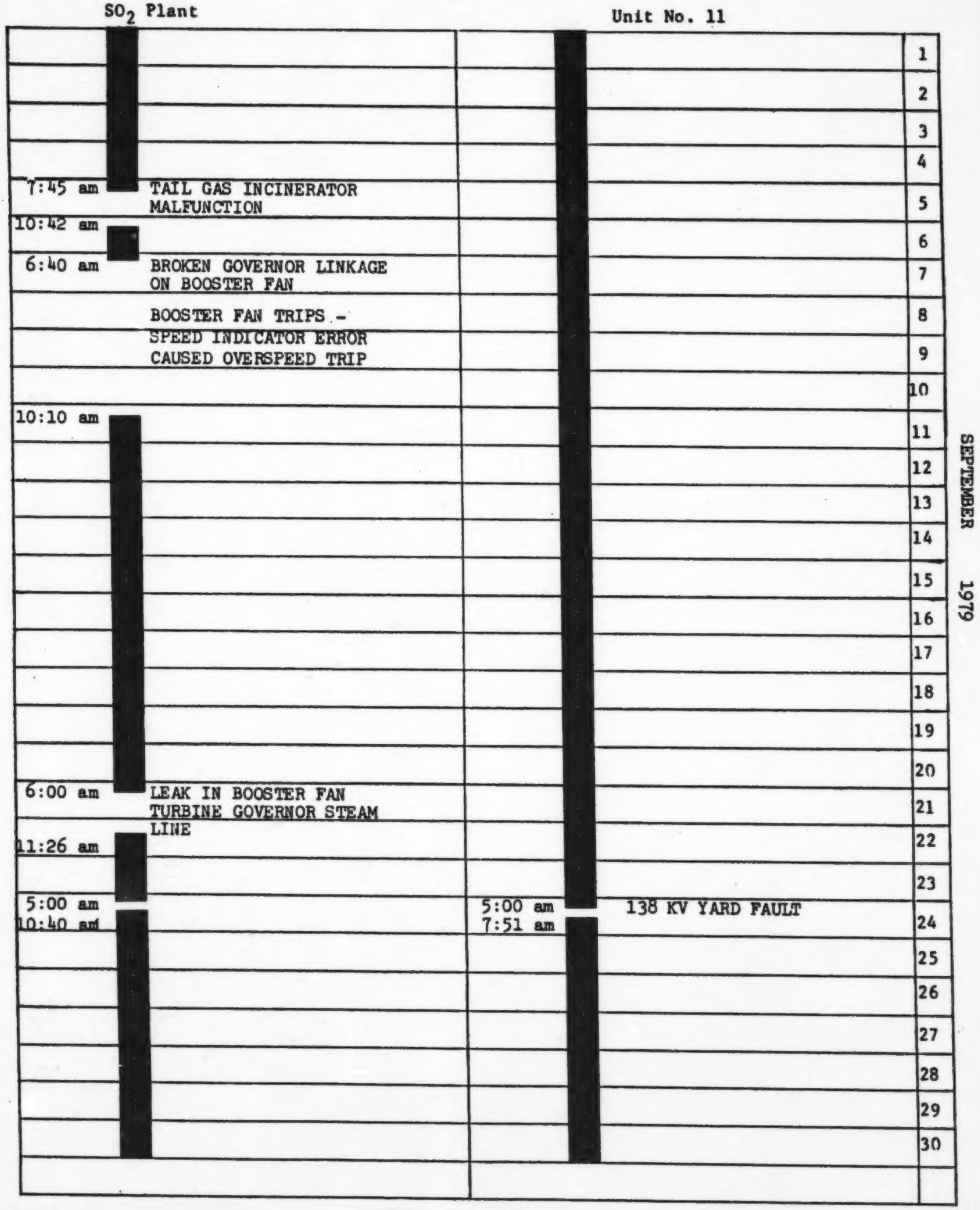
8:30 PM
11:11 PM
Lost Northeast Bus.

OPERATING HOURS

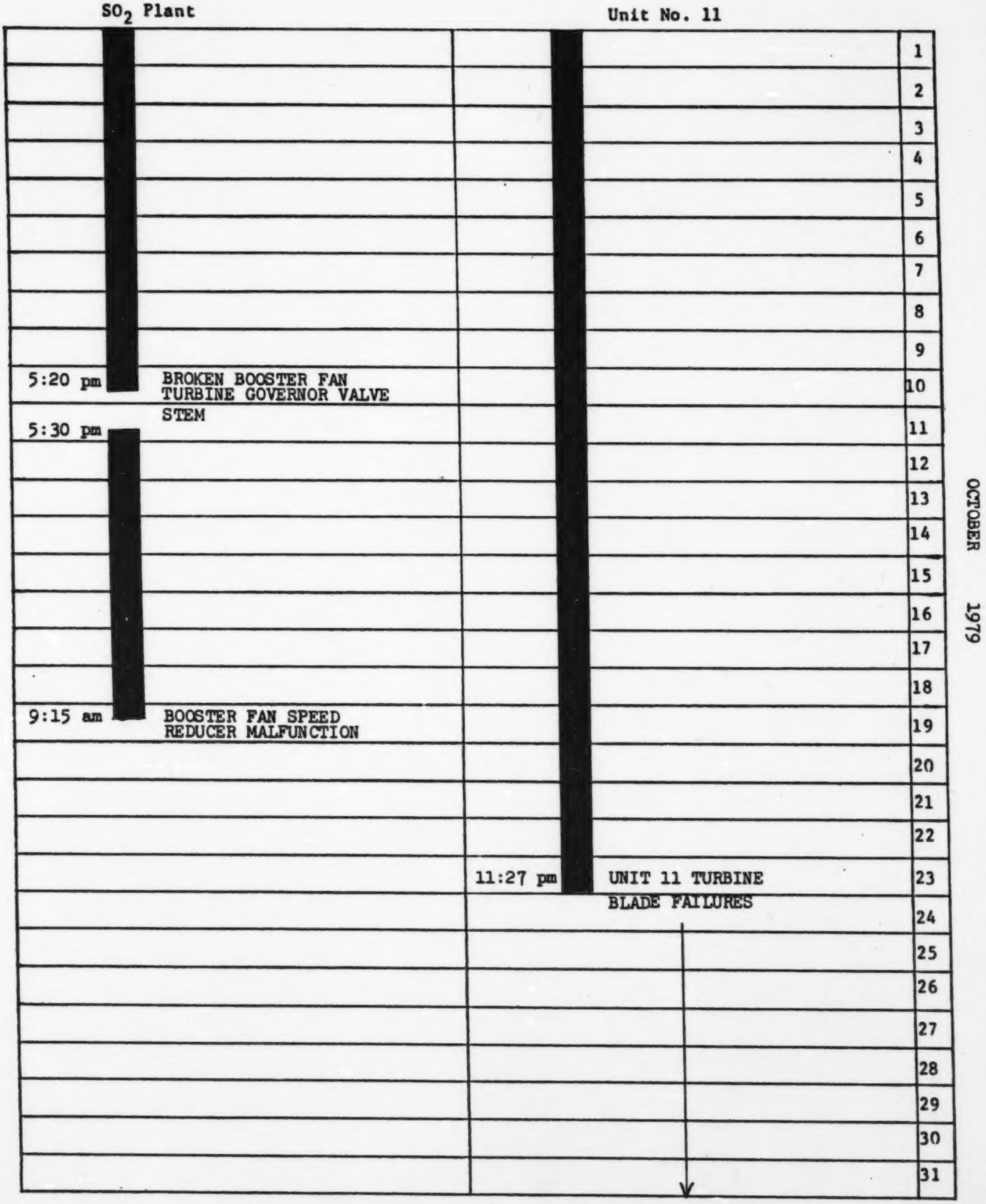


AUGUST 1974

OPERATING HOURS



OPERATING HOURS



OPERATING HOURS

SO_2 Plant	Unit No. 11	
CONTINUED SCHEDULED OUTAGE WITH UNIT 11	CONTINUED OUTAGE FOR TURBINE BLADE REPAIR AND SCHEDULED TURBINE AND BOILER MAINTENANCE	1
		2
		3
		4
		5
		6
		7
		8
		9
		10
		11
		12
		13
		14
		15
		16
		17
		18
		19
		20
		21
		22
		23
		24
		25
		26
		27
		28
		29
		30

NOVEMBER 1979

OPERATING HOURS

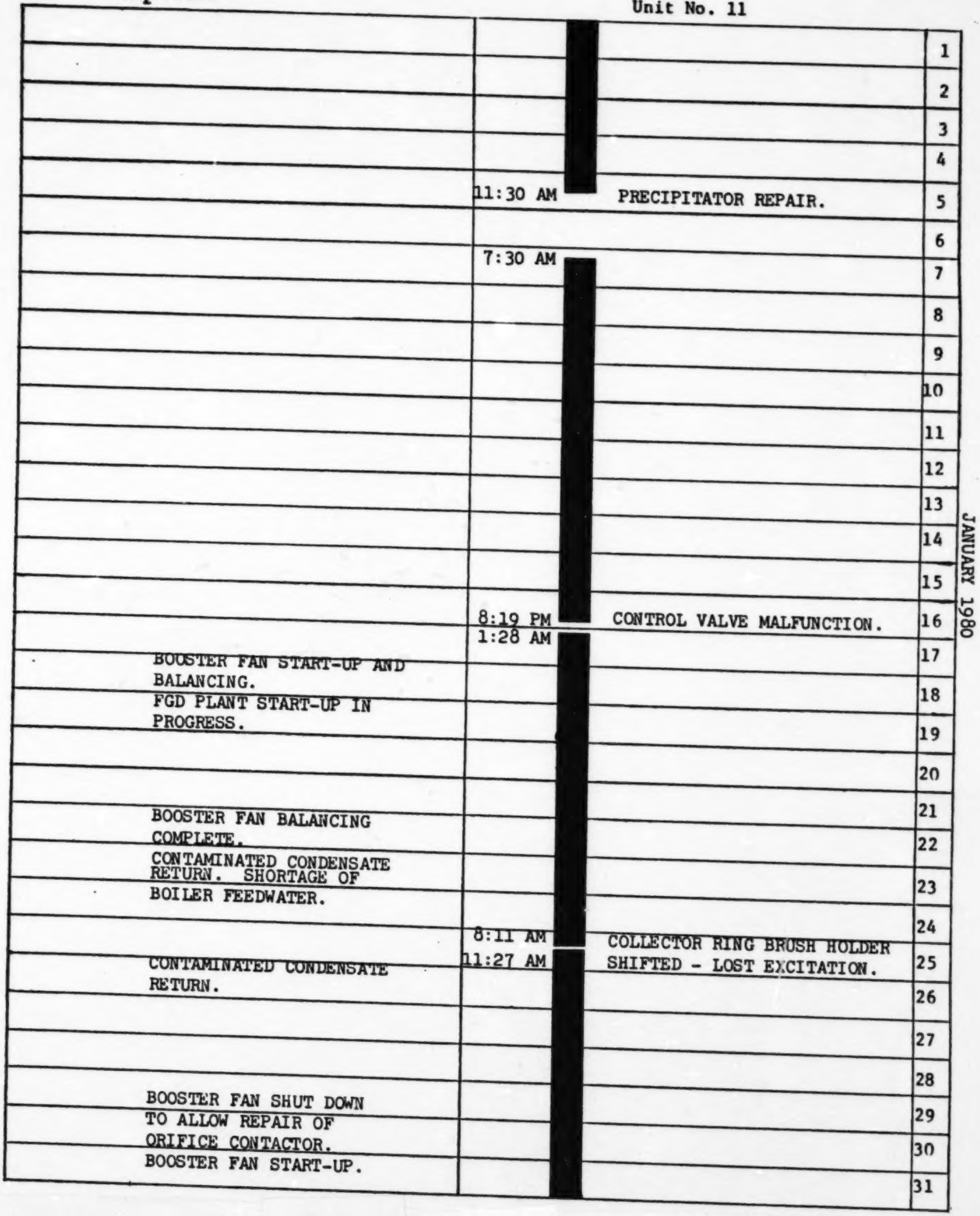
SO_2 Plant	Unit No. 11	
CONTINUED SCHEDULED	CONTINUED SCHEDULED	1
OUTAGE WITH UNIT 11.	OUTAGE FOR BOILER AND TURBINE MAINTENANCE.	2
		3
		4
		5
		6
		7
		8
		9
		10
		11
		12
		13
		14
		15
		16
		17
		18
		19
		20
		21
		22
		23
	12:10 PM — 12:17 PM UNIT START-UP	24
	2:53 PM — 3:21 PM 9:31 PM — 9:53 PM	25
	8:15 PM 11:30 PM	26
	8:35 PM	27
	12:13 AM	28
	6:45 PM HIGH EXHAUST HOOD TEMP. & VIBRATION.	29
	12:32 AM — 5:50 AM 12:50 AM TRIP CHECKS.	30
	12:40 AM — 6:30 AM 7:10 AM	31

DECEMBER 1979

OPERATING HOURS

SO_2 Plant

Unit No. 11



PRECIPITATOR REPAIR.

CONTROL VALVE MALFUNCTION.

BOOSTER FAN START-UP AND BALANCING.

FGD PLANT START-UP IN PROGRESS.

BOOSTER FAN BALANCING COMPLETE.

CONTAMINATED CONDENSATE RETURN. SHORTAGE OF BOILER FEEDWATER.

COLLECTOR RING BRUSH HOLDER SHIFTED - LOST EXCITATION.

CONTAMINATED CONDENSATE RETURN.

BOOSTER FAN SHUT DOWN TO ALLOW REPAIR OF ORIFICE CONTACTOR.

BOOSTER FAN START-UP.

NORTHERN INDIANA PUBLIC SERVICE COMPANY
FGD PLANT OPERATING HOURS

FEBRUARY, 1980

DATE	FGD PLANT		UNIT NO. 11		DATE
1		Orifice			1
2		Contactor			2
3		Repairs			3
4					4
5					5
6			Off 3:42pm On 4:45pm	Turbine Control Valve Malfunction	6
7					7
8	On 11:28am Off 1:45pm				8
9		Piping Leak In			9
10		Reduction Area			10
11	On 1:45pm				11
12	Off 10:25am	Leak In Shell Of "A"			12
13	8:00am	Regenerator			13
14		SO_2 Plant Available			14
15		But Condensate Quality			15
16		Problem in Generating			16
17		Station			17
18					18
19	On 2:03pm Off 4:15pm				19
20		Leak In Regenerator			20
21	On 9:48am				21
22	Off 8:12am On 10:29am	Leak In Regenerator			22
23					23
24					24
25	Off 10:40am On 2:00pm	Replace Leaking SO_2 S. Htr.			25
26					26
27	Off 9:15am	#11 To Come Off Line	Off 10:40pm	Generator H_2 Leak	27
28					28
29					29

NORTHERN INDIANA PUBLIC SERVICE COMPANY
FGD PLANT OPERATING HOURS

MARCH, 1980

DATE	FGD PLANT	UNIT NO. 11	DATE
1	SO$_2$ Compressor	█	1
2	Problems	█	2
3		█	3
4		█	4
5		█	5
6		█	6
7		█	7
8		█	8
9		█	9
10		█	10
11	On 0:10pm Off 2:20pm █ SO$_2$ Leak In Reduction Area Piping	█	11
12	On 0:45pm █	█	12
13	█	█	13
14	█	█	14
15	█	█	15
16	Off 4:22pm █ Incinerator Control	█	16
17	Failure	█	17
18	On 10:20am Off 12:01pm █ Unit 11 Scheduled To	Off 7:09pm █ Remove Turbine Stop	18
19	Be Off	Valve Screens And	19
20		Check H$_2$ Seals	20
21			21
22			22
23			23
24			24
25		On 1:02pm █	25
26		Off 10:56pm █ False Trip - Ground Protection Relay	26
27		On 6:22am █	27
28		█	28
29		█	29
30		█	30
31	On 11:45am Off 3:50pm █ Sulfur Blockage In Reduction Area	█	31

NORTHERN INDIANA PUBLIC SERVICE COMPANY
FGD PLANT OPERATING HOURS

APRIL, 1980

DATE	FGD PLANT	UNIT NO. 11	DATE
1	Sulfur Blockage	█	1
2	In Reduction Area	█	2
3	On 11:40am Off 1:40pm █ Sulfur Blockage	█	3
4	In "B" Claus Reactor	█	4
5		█	5
6		█	6
7		█	7
8		█	8
9		█	9
10		█	10
11		█	11
12	On 10:15am █	█	12
13	█	█	13
14	█	█	14
15	█	█	15
16	█	█	16
17	█	█	17
18	█	█	18
19	█	█	19
20	█	█	20
21	█	█	21
22	█	█	22
23	█	█	23
24	Off 1:40pm █ Unit 11 Off	Off 3:48pm █ Condenser Leaks	24
25		On 9:56am █	25
26	SO_2 Plant	█	26
27	Start Up	█	27
28	On 11:20am █	█	28
29	█	█	29
30	█	█	30

NORTHERN INDIANA PUBLIC SERVICE COMPANY
FGD PLANT OPERATING HOURS

MAY, 1980

DATE	FGD PLANT			UNIT NO. 11			DATE
1		■			■		1
2	Off 1:15pm	■			■		2
3			Repair		■		3
4			Bearing		■		4
5			Failure		■		5
6			In The		■		6
7			Evaporator	Off 9:36am On 12:01pm	■	False Trip - Generator Protection Relay	7
8			Pump Motor		■		8
9	1:15pm	—	SO_2 Plant Available		■		9
10	1:25pm	—	But No Steam From Unit 11		■		10
11			SO_2 Plant		■		11
12			Start Up		■		12
13	On 10:30am Off 1:45pm	■	Reduction Area SO_2		■		13
14			Leak & SO_2 Compressor		■		14
15			Turbine Governor		■		15
16			Steam Leak		■		16
17		—	SO_2 Plant	Off 2:32am		Dispatcher - Low	17
18			Available			System Load	18
19		—	SO_2 Plant Start Up	On 9:28am	■		19
20	On 1:05pm	■			■		20
21		■			■		21
22		■			■		22
23		■			■		23
24	Off 7:00pm	■			■		24
25			SO_2 Plant Available		■		25
26			Low System Load	Off 2:27am		Dispatcher - Low	26
27						System Load	27
28				On 9:27am Off 4:29pm	■	Precipitator Problem	28
29				1:40pm	—	Available	29
30						Dispatcher - Low	30
31						System Load	31

REFERENCES

1. R. C. Adams, T. E. Eggleston, J. L. Haslbeck, R. C. Jordon, and Ellen Pulaski. Demonstration of Wellman-Lord/Allied Chemical FGD Technology: Boiler Operating Characteristics, EPA Contract No. 68-02-0235 (1977).
2. R. C. Adams, S. J. Lutz, and S. W. Mulligan. Demonstration of Wellman-Lord/Allied Chemical FGD Technology: Acceptance Test Results, EPA Contract No. 68-02-1877 (1979).
3. F. A. Ayer. Proceedings: Symposium on Flue Gas Desulfurization - Las Vegas, Nevada, EPA Contract No. 68-02-2612 (1979).
4. R. C. Adams, J. Cotter, and S. W. Mulligan. Demonstration of Wellman-Lord/Allied Chemical FGD Technology: Demonstration Test First Year Results, EPA Contract No. 68-02-1877 (1979).

MAGNESIUM FGD AT TVA: PILOT AND FULL-SCALE DESIGNS

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ABSTRACT

This paper discusses pilot and full-scale magnesium flue gas desulfurization (FGD) designs by TVA.

The full-scale (600-MW equivalent) magnesium FGD design is for operation at high and low load factors for a high sulfur coal. After a process and system chemistry (magnesium sulfite/bisulfite) description, the paper describes the FGD equipment and system operation which includes an onsite acid plant. The second part of the paper discusses information on a test program and schedule of a pilot plant being considered by TVA to verify the magnesium FGD design.

MAGNESIUM FGD AT TVA: FULL-SCALE DESIGN

INTRODUCTION

TVA's involvement with magnesium flue gas desulfurization (FGD) systems has included some testing of MgO on the 1-MW cocurrent scrubber pilot plant at Colbert Steam Plant in 1976, a 10-MW cocurrent prototype system at Shawnee Steam Plant in 1978, and various economic studies.[1,2,3,4] The use of lime or limestone FGD systems requires large land areas for disposal of the calcium sludge produced. A regenerable FGD system eliminates the calcium sludge, provides a salable byproduct, and regenerates the absorbent for SO_2 removal. United Engineers and Constructors (UE&C) assisted TVA in the process design of the FGD system and provided detailed engineering support, especially in the magnesium regeneration area.

PROCESS DESCRIPTION/SYSTEM CHEMISTRY

The magnesium FGD design uses magnesium sulfite-bisulfite chemistry for SO_2 absorption, operating at a pH of approximately 6.0. After absorption of SO_2 by magnesium sulfite to form magnesium bisulfite, magnesium oxide is added to the absorber recycle tank to react with the bisulfite and precipitate magnesium sulfite. The sulfite can exist as tri- or hexa-hydrate depending on startup and operating conditions. The equipment is being sized to operate with either compound. In the trihydrate mode, the magnesium salts (10 percent solids) will be dewatered to 70 percent solids and then dried to remove all free water and most of the bound moisture (see Figure 1). The use of a sulfite storage silo provides the ability to operate the absorption and dewatering areas completely independent of the downstream regeneration area and acid plant. The calciner is designed to decompose the magnesium sulfite (with an average of 1/2 mole of bound water per mole of magnesium sulfite) and sulfate (with 7 mole of bound water per mole of magnesium sulfate) into magnesium oxide and SO_2. At a calciner operating temperatures of 1800°F, only 60 percent of the sulfate is decomposed. In actual operation the thermal decomposition of sulfate would be optimized to give the highest necessary percentage decomposition at the lowest temperature. The offgas from the calciner is designed for an SO_2 concentration of 17 volume percent for feed to a single contact acid plant. Since the acid plant tail gas goes back to the main plenum, the SO_2 emissions from the acid plant are part of the overall plant SO_2 emissions.

Both the Philadelphia Electric Company's (PECo) experience at Eddystone and the brief TVA experience at Shawnee indicates that there is little, if any, solid magnesium sulfate formed as a solid in the absorber recirculating material. The dissolved magnesium sulfate level in the absorbing slurry would be about 30 percent (by weight) using only the liquor in the trihydrate centrifuge cake to the dryer as a purge stream. However, this 30 percent is based on 10 percent oxidation of the absorbed SO_2 to sulfate, 60 percent decomposition of sulfate

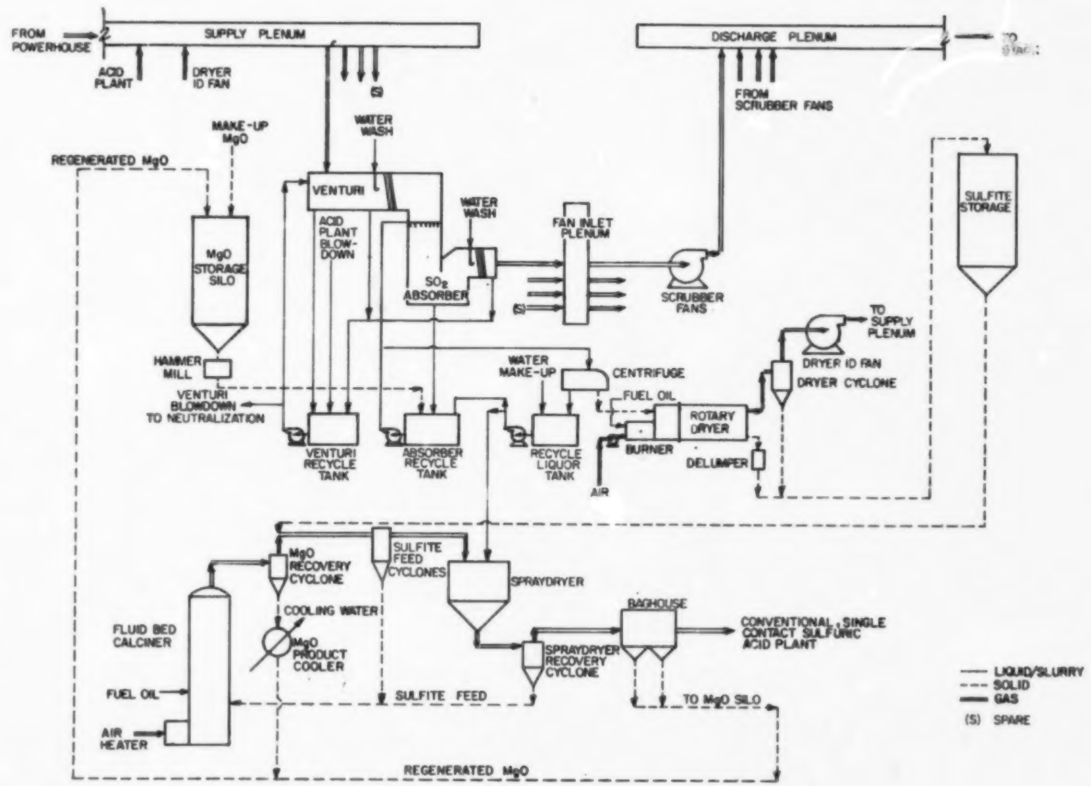
FROM POWERHOUSE
SUPPLY PLENUM
ACID PLANT
DRYER ID FAN
(S)
DISCHARGE PLENUM
TO STACK
FROM SCRUBBER FANS
REGENERATED MgO
MAKE-UP MgO
WATER WASH
VENTURI
ACID PLANT BLOW-DOWN
MgO STORAGE SILO
WATER WASH
FAN INLET PLENUM
SO$_2$ ABSORBER
(S)
SCRUBBER FANS
SULFITE STORAGE
TO SUPPLY PLENUM
HAMMER MILL
WATER MAKE-UP
CENTRIFUGE
DRYER ID FAN
DRYER CYCLONE
VENTURI BLOWDOWN TO NEUTRALIZATION
FUEL OIL
ROTARY DRYER
VENTURI RECYCLE TANK
ABSORBER RECYCLE TANK
RECYCLE LIQUOR TANK
BURNER
DELUMPER
AIR
MgO RECOVERY CYCLONE
SULFITE FEED CYCLONES
SPRAYDRYER
COOLING WATER
BAGHOUSE
FLUID BED CALCINER
MgO PRODUCT COOLER
CONVENTIONAL, SINGLE CONTACT SULFURIC ACID PLANT
SPRAYDRYER RECOVERY CYCLONE
FUEL OIL
SULFITE FEED
TO MgO SILO
AIR HEATER
REGENERATED MgO
LIQUID/SLURRY
SOLID
GAS
(S) SPARE

Figure 1: FGD System Process Flow Diagram

in the calciner, and a trihydrate mode of operation (70 percent solids in the centrifuge cake). The PECo experience shows less than 10 percent oxidation but at a higher operating pH in the absorber. Operation at the lower pH may produce more oxidation and a hexahydrate crystal reducing the amount of liquor evaporated in the rotary dryer. The hexahydrate cake is easier to dewater than the trihydrate. This operation would raise the dissolved sulfate level in the slurry to well-in-excess of 30 percent by weight.

In order to provide flexibility of operation and for thermal protection of the baghouse used for final magnesium particulate collection before the acid plant, a spray dryer was placed between the calciner and the equipment used for final magnesium particulate removal. The spray dryer will use centrate from the centrifuge containing dissolved magnesium sulfite, bisulfite, and sulfate. In this manner, magnesium value in the form of $MgSO_3$ and $MgSO_4$ can be recovered. This sulfite/sulfate mixture will be recycled to the sulfite silo for feed to the calciner. Since there is a 60 percent sulfate decomposition in the calciner, this use of the spray dryer allows for control of the dissolved sulfate in the absorbing slurry, depending on actual operating conditions. For the design case, the sulfate level will be 24 percent $MgSO_4$ (by weight).

FGD SYSTEM EQUIPMENT

In the absorption area, particulate, chlorides, and SO_2 are removed by four (4) venturi-type prewash and absorber modules. The venturi with its own two-stage mist eliminator removes chlorides, SO_3, and fly ash in a separate liquor loop. Any chlorides which are not removed in the venturi loop will form magnesium chloride in the absorber loop and could cause corrosion problems in the regeneration area and acid plant feed gas clean-up system. The blowdown from the venturi recycle tank is neutralized in a separate facility with hydrated lime and is then pumped to the plant disposal area. The venturi liquor loop is not neutralized and, therefore, operates at a pH of less than one. In addition to the SO_3 and chlorides removed in the venturi, additional dilute sulfuric waste acid from the acid plant feed gas humidification/cooling towers and wet ESP's are added to the venturi recycle tank. This acid blowdown from the acid plant feed gas clean-up system results from the water quenching of the calciner offgas which contains about 0.6 volume percent SO_3.

To raise the venturi liquor pH to about 1.0, consideration is being given in the design to pumping the acid plant blowdown directly to the neutralization facility or to the absorber recycle liquor tank. In theory, the waste acid stream would form more magnesium bisulfite in the recycled liquor and require more magnesium oxide in the absorber recycle tanks. This would increase the load on the regeneration area and acid plant but increase the amount of sulfuric acid recovered and decrease the lime required for neutralization. Of the anticipated lime requirement for neutralization, 50 percent is required for the acid plant feed gas clean-up system blowdown. In either case, the low pH and high chloride concentration will require corrosion protection of the equipment. For the venturi-type prescrubber, besides Inconel 625 for high abrasion areas and FRP internals, the mist eliminators could be factory-installed on special Inconel strips to avoid field installation of the blades. An organic lining could be used to protect the carbon steel venturi shell. Efforts would have to be made to ensure that there is minimal interior work after lining installation (such as factory-installed mist eliminator blades).

The SO_2 absorbers and their two-stage mist eliminators are designed for operation in a cocurrent gas and slurry mode at 4.6 meters per second (15 feet per second) and an L/G of 4.5 liters per actual cubic meter (34 gallons/1000 acf), and for mist elimination with horizontal gas flows of 6.1 meters per second (20 feet per second). The cocurrent absorbers are based on TVA's earlier test work with cocurrent absorbers at Colbert (1 MW) and Shawnee (10 MW). Gas/slurry flows in the absorber/mist eliminator, especially the wet elbow (180° turn) where the bulk of the entrained slurry is removed (see Figure 2), can cause large variations in the gas velocity profile and produce solids deposition on the mist eliminator. The absorbers could be 316L stainless steel; a need for higher grades of alloys is not anticipated due to the low concentration of chlorides projected for the absorber slurry.

There is no reheater in this design since saturated flue gas can be mixed with unscrubbed flue gas after the scrubber for the flue gas reheat. The scrubber fans have Inconel 625 rotors and 316L housings and the bypass fans, if needed, would be Corten construction. The lining for this very corrosive flue gas would require extensive corrosion resistance due to the mixture of water vapor (scrubbed gas) and SO_3/chlorides (bypassed gas), below the average boiler flue gas acid dew point.

The absorber recycle pumps and tanks are designed for normal slurry service, similar to a limestone FGD system except for the use of a waterless seal to assist in water balance maintenance. Since the liquor in the centrifuge cake (going to the dryer for evaporation) is only 35 GPM and any seal water for these large 10,000 GPM recycle pumps would be at least 10 to 15 GPM per pump (for a total of 60-90 GPM for the six (6) operating pumps), the use of seal water for the slurry would require a purge of magnesium liquor to maintain the absorber loop water balance. With all evaporation of water for quenching the flue gas taking place in the venturi portion of the process, the major water loss from the absorber-regeneration section is through the centrifuge cake into the drying system. Since each GPM lost to blowdown to allow for water in-leakage from the pump seals was worth $200,000 in makeup magensia value over the life of the plant, TVA designed for pumps that did not require seal water. These types of pumps had been used successfully at Philadelphia Electric Company (PECo) to help maintain a water balance.

The regeneration area required several difficult decisions. Since the calciner offgas is at 1800°F, economic considerations dictated some heat be recovered. Based on PECo's experience, the original design of a shell and tube heat exchanger to preheat fludizing air would present major operating difficultues caused by the fouling of tube side heat transfer surfaces. The

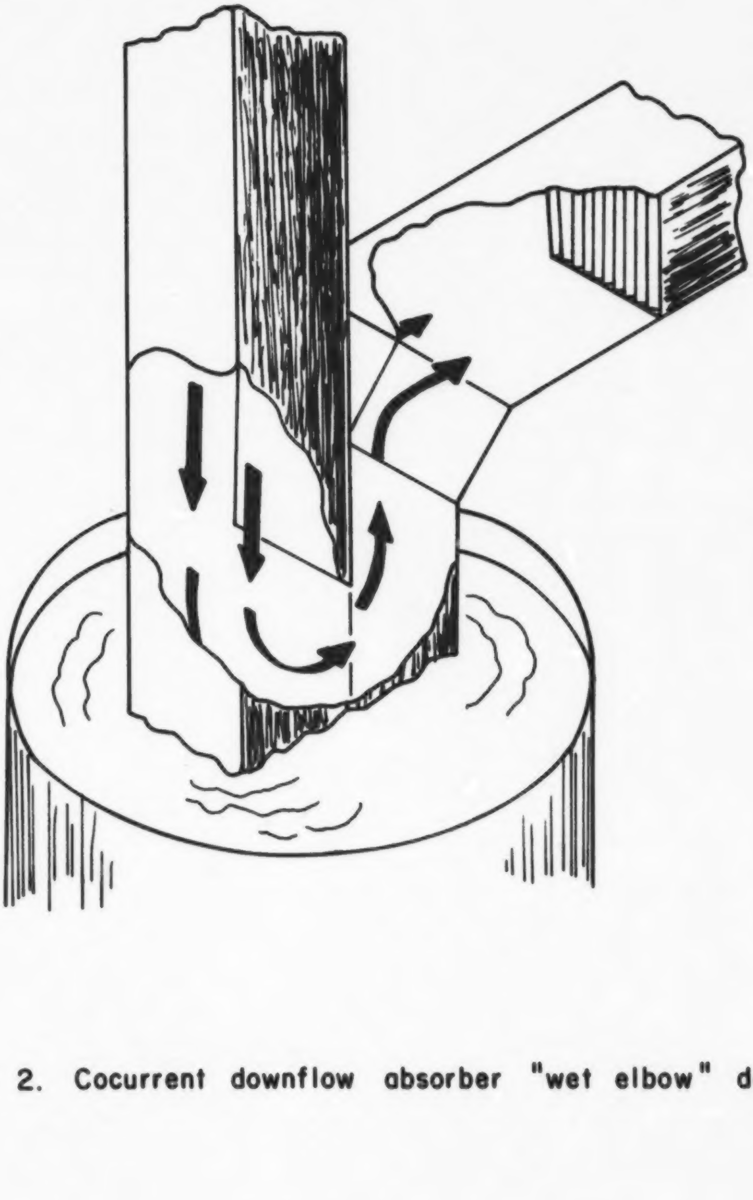

Figure 2. Cocurrent downflow absorber "wet elbow" design.

SO_2-rich offgas contains traces of SO_3 and the latter apparently recombines with the regenerated magnesium oxide to form a solid sulfate. Although the TVA design has MgO product recovery cyclones which are more efficient than those at PECo, there is some indication that the recombination is SO_3 controlled, or a diffusion limited reaction. Therefore, no matter how efficient the product recovery cyclones, there will always be recombination and potential subsequent pluggage. The method chosen to avoid this problem in the design is to preheat the magnesium sulfite solids being fed to the calciner in a manner which can be called "suspension heating." The sulfite solids at ambient temperature from the silo are injected into the 1800°F calciner offgas after the recovery cyclones. The sulfite solids are preheated and then recovered in the sulfite cylones. Although the inlets of these high efficiency cyclones are small, very little pluggage would be expected due to the high gas velocities in the cyclones. This type of solids preheating is now used in the cement industry and its successful operation in magnesium FGD requires that sulfite solids not be heated to the decomposition temperature during suspension heating.

The other difficult problem in the regeneration area concerned the use of either an electrostatic precipitator (ESP) or a fabric filter for final magnesium oxide particulate removal before the offgas passed to the acid plant. Any of this particulate which reaches the acid plant is lost from the regeneration loop and may poison the acid plant catalyst if not removed by the humidification or gas cooling towers of the acid plant feed gas clean-up system. The TVA design required 99 percent removal of the particulate remaining after the three sets of upstream cyclones. Originally for the design concept it was desirable for the final MgO particulate collection device to operate at temperatures in excess of 600°F, collect greater than 98 percent of the MgO and provide reliable operation. At first an ESP was selected over a fabric filter. The fabric filter's maximum operation is about 500°F; therefore, it is not suitable for this application. However, this selection was before the addition of a spray dryer for sulfate control to the regeneration area and before investigations were carried out on ESP collection of magnesium oxide. The change from an ESP to a fabric filter was finally resolved based on the operating characteristics of ESP collection of magnesium oxide and the realization that once the spray dryer was added, there was no reason to worry about high temperature excursions (greater than 500°F) affecting fabric material. PECo operated a pilot ESP on the regeneration offgas and verified the information that had been accumulating from ESP manufacturers and magnesium oxide supliers. Effective ESP operation on magnesium oxide requires temperatures in excess of 600°F and would be preferable at 700°F. In other systems this may have been acceptable but the proper operation of the spray dryers requires the largest temperature differential possible for maximum flexibility in controlling sulfate concentrations. A fabric filter would allow operation at the original design temperature of 450°F for final particulate removal.

FGD SYSTEM OPERATION

The magnesium FGD system design has an onsite acid plant containing two trains and with a total capacity of 350 tons per day. There are four 750,000-gallon storage tanks which can provide up to 60 days of storage. The acid plant can produce either 93 or 98 percent acid, depending on market conditions. The key to the system operation is the fact that storage silos for magnesium oxide and magnesium sulfite allow decoupled operation of the regeneration area/acid plant from that of the power plant. The absorber and dewatering/dryer areas are sized to treat the required flue gas at full load. Although the future yearly capacity factor for this FGD system would be significantly less than 60 percent, these systems must be sized to accommodate full load operation during peak periods.

The regeneration area and acid plant are not designed to follow load as are the absorbers (three operating and one spare), the centrifuges (four), and the rotary dryers (two). The regeneration area is gas flow dependent and the acid plant is SO_2 dependent. Therefore, a sulfite storage silo will be located after the rotary dryer. Correspondingly, there is a magnesium oxide storage silo for the regenerated magnesium after the recovery cyclones. Since the seasonal electric loads will probably not correspond with the seasonal industrial acid demand, typical operation will probably be as follows: high load on the absorbers (filling the sulfite silo) with low load on the regeneration/ acid plant (drawing down the oxide silo); or low absorber load (drawing down the sulfite silo) with high load on the regeneration/acid plant (filling the oxide silo).

The second section of this paper provides information on a test program and schedule of a pilot plant being considered by TVA. The research and development areas of concern are identified for each of the major process steps. A number of research and development tasks which would provide the base technology for the successful operation of the full-scale magnesia FGD system are discussed.

MAGNESIUM FGD AT TVA: PILOT DESIGN

INTRODUCTION

As mentioned earlier, land availability around existing and future fossil fuel steam plants could limit the disposal of wastes for any FGD system that has a "throwaway" product. However, it is generally agreed that magnesium FGD technology is not as well developed, particularly insofar as the regeneration portion of the process is concerned, as that of lime/limestone scrubbing.

PILOT OBJECTIVES

The operation of the pilot plant will simulate the full-scale TVA magnesium FGD design in every practical way for the purposes of this study. Primary objectives of the pilot plant program are as follows:

1. Develop valuable in-house experience with the MgO process.
2. Anticipate potential problems with the TVA MgO process chemistry or equipment.
3. Evaluate the long-term effects of process contaminants such as fly ash, chlorides, and trace elements which evolve from the burning of coal by reliability runs of several months' duration.
4. Study the formation and properties of magnesium sulfite hexahydrate versus trihydrate crystals with respect to potential solids handling problems.
5. Study the regenerated MgO absorption capability after several cycles.

Secondary objectives are to develop process design improvements and innovations such as sulfur production, and the use of coal for drying and calcining through additional studies not yet completely defined. All these studies will be done in conjunction with EPA. The secondary objectives will be met in such a way as not to impact the schedule for fulfillment of primary objectives. Most of these objectives are process related and have general application to magnesium FGD technology.

TVA has retained the magnesium FGD design architect/engineer, United Engineers and Constructors, Inc., and had the magnesium FGD design scaled down to the 1-MW level. The pilot plant would not include the acid production facilities.

Sections of the pilot plant (primarily the regeneration and drying sections) will be "skid" mounted to provide maximum flexibility in operation. Skid mounting will allow the operation of the pilot plant to be broken down into a Phase I and Phase II operational scheme. Phase I operation will be chiefly concerned with the primary objectives of the pilot plant as previously listed. Phase II operation will be concerned with additional studies.

These additional studies will develop process design improvements and innovative concepts. Examples of such improvements and innovative concepts may include but not be limited to the following:

1. Using coal instead of oil in the drying and calcining of the magnesium sulfite.
2. Spray drying for SO_2 absorption instead of wet scrubbing.
3. Testing of a sulfur-producing technology using coal as a reductant instead of natural gas or oil.

The current schedule calls for the pilot plant to start up in the fall of 1981, with the intention of providing operational experience and some solutions to the problems mentioned earlier.

RESEARCH AND DEVELOPMENT AREAS OF CONCERN

Identified in the following subsections for each of the major processing steps are more details of a number of research and development tasks (outlined above) which would provide base technology for the successful operation of a full-scale magnesium FGD system.

PRESCRUBBER

The pH of the prescrubber solution has been calculated to be lower than 1.0 due to dissolved HCl and H_2SO_4. In the particulate scrubber, the hot flue gas is contacted with a slurry of fly ash and river water. Most of the particulate (fly ash) and hydrogen chloride, and a variable fraction of the sulfur trioxide, are removed in the scrubber liquor. The blowdown slurry from the particulate scrubber, therefore, is acidic and the water may contain high concentrations of dissolved solids, trace metals, toxic organics, and radionuclides either leached out from the fly ash or absorbed from the flue gas.

In this acidic environment only expensive alloys such as Inconel 626 or Hastelloy G, or an organic lined metal alloy, may be suitable. Should the prescrubber be made of materials such as 316L stainless steel, addition of a neutralizing agent such as caustic soda or lime/limestone would be necessary to raise the pH to at least 3.0. This addition would be undesirable since it introduces additional dissolved solids into the system and may complicate the disposal of the prescrubber blowdown as described later. Various materials of construction (through the use of coupons) will be tested in the pilot plant venturi-type prescrubber to identify those alloys and organic lined metals able to withstand the low pH environment without neutralization to a higher pH.

SO_2 ABSORPTION

For this portion of the FGD process there remain two concerns: (1) the chemical effects of chlorides, fly ash, and dissolved magnesium salts on sulfite oxidation and (2) magnesium sulfite trihydrate ($MgSO_3 \cdot 3H_2O$) vs. hexahydrate ($MgSO_3 \cdot 6H_2O$) formation. Experience has shown that the trihydrate crystals can be obtained during steady-state operation, but during startup, shutdown, or during nonsteady-state operation hexahydrate crystals have been observed. The crystals of the two hydrates have widely different handling properties and this fact can introduce difficulties in the solids separation and drying steps of the process. The operating parameters affecting the above items will be investigated.

As an alternative to wet scrubbing TVA is considering spray drying and subsequent fabric filter collection for SO_2 absorption as part of the Phase II operational scheme. Although definite plans have not been formulated, a spray dry/fabric filter would be investigated to determine optimum operating conditions. The regeneration section from the Phase I tests would be used to decompose the $MgSO_3$ to MgO and SO_2. This FGD system would have the advantages over the wet scrubbing system of no flue gas reheat and elimination of the drying step.

DRYING

Two control schemes described below will be investigated to determine optimum operating parameters with regards to (1) sulfite to sulfate oxidation during drying, (2) $MgSO_3$ decomposition, and (3) economical operating conditions.

With the first scheme, two variables--the airflow (primary combustion air and dilution air) and fuel flow--will be controlled in series to maintain a constant dryer gas discharge temperature. Fuel flow control follows combustion chamber temperature, thereby increasing lag time for response and decreasing control accuracy. The dryer discharge gas temperature control point senses the gas temperature and demands an inverse gas flow change before a fuel adjustment has fully responded to its initial demand. Also, the refractory lined combustion chamber is an excellent heat sump which further increases lag time for fuel adjustment.

For the second control scheme, common practice in other industries is to maintain constant gas flow at a maximum rate that has an acceptable dew point and carry-over dust load; the discharge gas temperature then controls the fuel flow directly. Such a control method will eliminate one of the two control variables and may reduce control lag time to a minimum.

REGENERATION

There are two concerns about the fate and possible accumulation of minor elements such as calcium, sodium, and chlorine in the regeneration portion of the FGD process: (1) Does the chlorine leave the process perhaps as HCl, or does it remain in the process tying up magnesium salts, and (2) Does a molten liquid phase of $MgCl_2$, NaCl, or $CaCl_2$ exist in the calciner offgas?

A predicted composition of the calciner feed stream for the magnesium FGD design is summarized in Table 1.[5] At the desired operating temperature (1800°F) the output from the calciner was thermodynamically calculated and is shown in Table 2.[6] Thermodynamically, at 1800°F the decomposition of $MgSO_4$ to MgO, SO_3, SO_2, and O_2 is favorable; calcium is present in the solid phase as $CaSO_4$, and a liquid sodium sulfate/chloride melt is predicted. Most of the chlorine in the calciner feed leaves the system in the gas phase as HCl.

TABLE 1. CALCINER SOLIDS FEED

Compounds		Weight Percent
$MgSO_3$		70.0
$MgSO_4$		8.9
$MgCl_2$		9.8
Na_2SO_4		0.2
$CaCl_2$		0.4
Ash		10.6
	Total	99.9

Experience with PECo's prototype calciner (actually an entrained-bed reactor rather than a traditional fluid-bed reactor) at its Essex Chemical test facility has revealed that approximately 30 percent of the $MgSO_3$ feed forms hard, chemically unreactive MgO pellets. These pellets do not discharge with the MgO fines overhead in the offgas but rather eventually fill the calciner bed and have to be ground before reuse in SO_2 sorption. Examination of the high density MgO pellets indicates a possible double salt, $Mg_3 Ca(SO_4)_4$, as the root cause of pellet formation.[7] The MgO fines generated overhead are very fine (3-30 microns) and difficult to handle, thereby causing transportation and storage problems.

As discussed earlier, in the full-scale design fouling of the shell-and-tube air preheater exchanger by $MgSO_4$ was also experienced at PECo's regeneration test facility at Essex Chemical. The plugging was rapid and rendered the heat exchanger unusable. The magnesium sulfate is thought to form in the calciner offgas by recombination of MgO and the relatively small amounts of highly reactive SO_3. The recombination should be avoided not only because of heat exchanger plugging, but also because it introduces a recirculating load of $MgSO_4$ in the system that is inactive for SO_2 sorption.

The regeneration study will determine whether the predicted sodium sulfate/chloride melt and $CaSO_4$ in the gas phase are the cause of the problems at Essex Chemical and, if so, will determine the most economical solution to the shell and tube preheater fouling. Specifically, we propose to examine, as a function of the operating parameters (1) $MgSO_3$ and $MgSO_4$ decomposition, (2) MgO pellet formation, (3) $MgSO_4$ recombination, (4) optimum product yield, and (5) chlorine composition and purge rate.

TABLE 2. OUTLET CALCINER GAS COMPOSITION AT 1800°F

	Mole Percent
Gases	
CO_2	9.4
H_2O	5.5
N_2	53.7
NO	$8.75\ X\ 10^{-3}$
HCl	3.4
Cl^-	$1.40\ X\ 10^{-2}$
Cl_2	$2.11\ X\ 10^{-2}$
SO_2	11.6
SO_3	0.3
$MgCl_2$	$1.60\ X\ 10^{-3}$
O_2	1.7
Liquids	
Na_2SO_4	$1.16\ X\ 10^{-2}$
$MgCl_2$	$1.26\ X\ 10^{-5}$
NaCl	$3.75\ X\ 10^{-3}$
Solids	
MgO	13.6
$MgSO_4$	-
$CaSO_4$	0.7
	100.0

DESIGN IMPROVEMENTS AND INNOVATIVE CONCEPTS

In addition to spray drying and subsequent fabric filter collection for SO_2 collection, TVA is considering the following: (1) the use of coal or a coal-oil mixture in place of oil in the drying and calcining of the magnesium sulfite and (2) the direct production of elemental sulfur from the decomposition of $MgSO_3$. Although no definite plans have been formulated, minor modifications to the Phase I drying and regeneration equipment is all that is necessary to test the coal-fired option (excepting, of course, the addition of coal handling facilities). For the production of elemental sulfur, TVA has obtained the services of P. S. Lowell to expand upon an earlier EPA study on this topic.[8]

SUMMARY

Thus, we have identified proposed research and development studies on a number of potential problems. These problems, it is felt, are soluble with current state-of-the-art engineering knowledge. The magnesium FGD design offers promise as a technically viable, economically feasible process for recovering SO_2 as a useful product--sulfuric acid.

REFERENCES

1. Robards, R. F.; Cole, R. M.; Morasky, T. M.; "TVA's Cocurrent Scrubber Evaluation," Winter Annual Meeting of the Air Pollution Control Division of ASME, Atlanta, Georgia, 1977, 77-WA/APC-8.

2. Robards, R. F.; Moore, N. D.; Kelso, T. M.; and Cole, R. M.; "Cocurrent Scrubber Evaluation: TVA's Colbert Lime-Limestone Wet Scrubbing Pilot Plant," EPRI FP-941, Research Project 537-1; January 1979.

3. Marcus, E. G., and Wells, W. L., "Magnesium Oxide Testing on the 10-MW Cocurrent Scrubber at the Shawnee Steam Plant," in press.

4. McGlamery, G. G.; Torstrick, R. L.; Simpson, J. P.; and Phillips, Jr., J. F.; Conceptual Design and Cost Study; Sulfur Oxide Removal from Power Plant Stack Gas, EPA-R2-73-244, May 1973.

5. Lowell, P. S., "Technical Memorandum 009-02-02, Equilibrium Program and Thermodynamic Data Base for TVA Reducing Calciner," May 1980.

6. Lowell, P. S., "Technical Memorandum 009-02-08, Thermodynamic Analysis of the TVA $MgSO_3$ Colbert Pilot Calciner," May 1980.

7. Kelmer, A. D., "Magnesia Flue Gas Desulfurization--Status of Development and Needed Technological Support, Letter from Charles D. Scott (Assoicate Director, Chemical Technology Division, ORNL) to J. Frederick Weinhold (Director, Division of Energy Demonstrations and Technology, TVA), February 21, 1980.

8. Lowell, P. S.; Corbett, W. E.; Brown, G. D.; and Wilde, K. A.; Feasibility of Producing Elemental Sulfur from Magnesium Sulfite, EPA-600/7-76-030, October 1976.

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